THE ECONOMICS OF RESIDUAL FUEL OIL DESULFURIZATION

A Study for the

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Division of Air Pollution
Public Health Service

U. S. DEPARTMENT OF
HEALTH, EDUCATION, AND WELFARE

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FOREWORD

Under the Clean Air Act (Public Law 88-206, 88th Congress) the U. S. Department of Health, Education, and Welfare is charged with the responsibility of expediting research into the desulfurization of fuels. The Act provides for the letting of contracts to complement research activities performed directly by the Public Health Service of the Department and by other organizations under research grants.

Under contract to the Public Health Service, the Bechtel Corporation has investigated the costs of reducing the sulfur content of certain residual fuel oils. The Division of Air Pollution, Public Health Service, is publishing this summary report prepared by the Bechtel Corporation to make these research results available to all individuals and organizations who may find them useful.

THE ECONOMICS OF
RESIDUAL FUEL OIL DESULFURIZATION

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OBJECT AND SCOPE

No. 6 Residual Fuel Oils marketed in the United States contain an average sulfur content of about 1.6 percent, with a range encompassing a minimum of about 0.4 percent to a maximum of about 4.5 per cent. The emission of sulfur dioxide from furnaces burning such fuel is recognized as a cause of significant air pollution.

The objective of this study is to establish costs for reducing the sulfur content of fuel oils manufactured from high sulfur crudes. The desired maximum sulfur content for an acceptable fuel oil has been taken to be 0.5 wt. percent for this study.

Presented are eight case studies developed in detail with the aid of a Linear Programming (LP) model of a typical 100,000 barrel per day modern refinery. These eight cases are as follows:

Case No.	Description
1	The base case - a nominal 100,000 BPSD refinery which, with no restrictions on sulfur, would produce a residual fuel containing 1.63 percent of sulfur. Huntington Beach (California) crude is the feed stock. With respect to residual fuel sulfur content, the product is typical of U. S. production. Progressively severe restrictions are placed on sulfur content of residual fuel and the facilities required to meet these restrictions most economically are defined.
2	Case No. 1 with coking excluded. This forces utilization of the more expensive residual oil desulfurization process.
3	Case No. 1 with both coking and hydro-pretreatment of catalytic cracker feed stock excluded.
4	Case No. 1 with fuel oil production reduced to 10,000 BPSD.
5	In this case, the residual fuel sulfur content is held constant at 0.5 percent while the production rate is varied. All processing options are permitted.

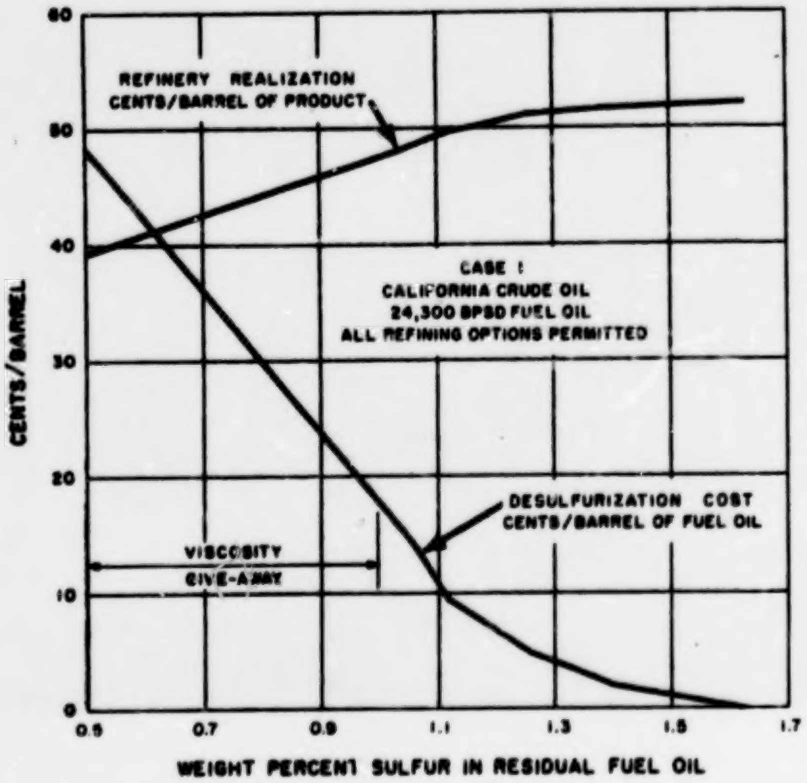
Case No.	Description
6	Case No. 1, in which 5,000 BPSD of Kuwait Crude Oil, and 500 BPSD of "imported" Kuwait No. 6 residual fuel oil are included in the refinery feed. As in Case No. 1, progressively severer restrictions are applied to the sulfur content of the residual fuel product.
7	The same as Case No. 6, but applied to an existing refinery optimized for operations without restriction on the fuel oil sulfur content. It is assumed that the "existing" refinery is fully depreciated, so that capital charges are assigned only on new units required to meet the restrictions on fuel oil sulfur.
8	A special case, applicable to a Gulf Coast refinery, where cheap natural gas fuel is available for potential conversion to hydrogen.

SUMMARY

The results of the eight case studies are presented briefly in this summary. Later in the report each of these cases is described in detail with all of the important computer results presented in tables. Except for Case No. 5 which deals with varying fuel oil production rates, all of the eight cases are concerned with the increase in refining costs as progressively stringent sulfur restrictions are applied. Desulfurization costs shown are calculated as differential costs relative to operation for that case where there is no sulfur restriction imposed. Since the quantity of the other refinery products and their quality specifications are held constant for all cases, these relative costs are realistic indications of the effect of a sulfur restriction. The various cases may be compared with each other by the refinery realization curves. These represent a relative profit which is calculated by subtracting variable costs from the value of the products. Since only variable costs are considered, the magnitudes of these numbers have no significance except in relation to each other.

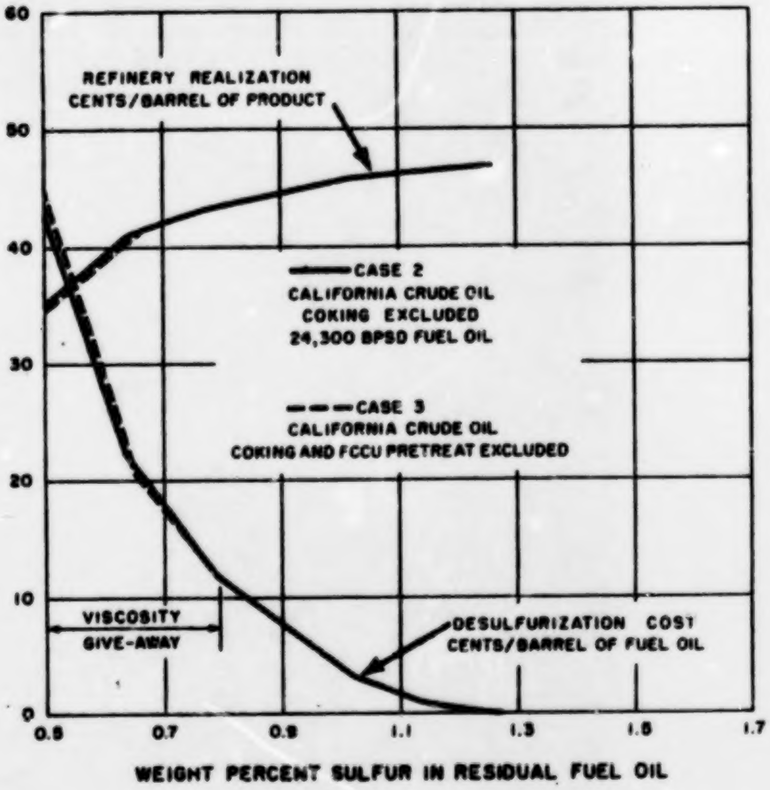
The area of the chart marked "viscosity give-away" indicates the range in which the computer could not economically meet both the viscosity and sulfur specifications for the residual fuel oil. In these cases the product is more fluid than that required by the market. For some applications, this low viscosity fuel oil could command a small premium, but no such credit was taken in this study.

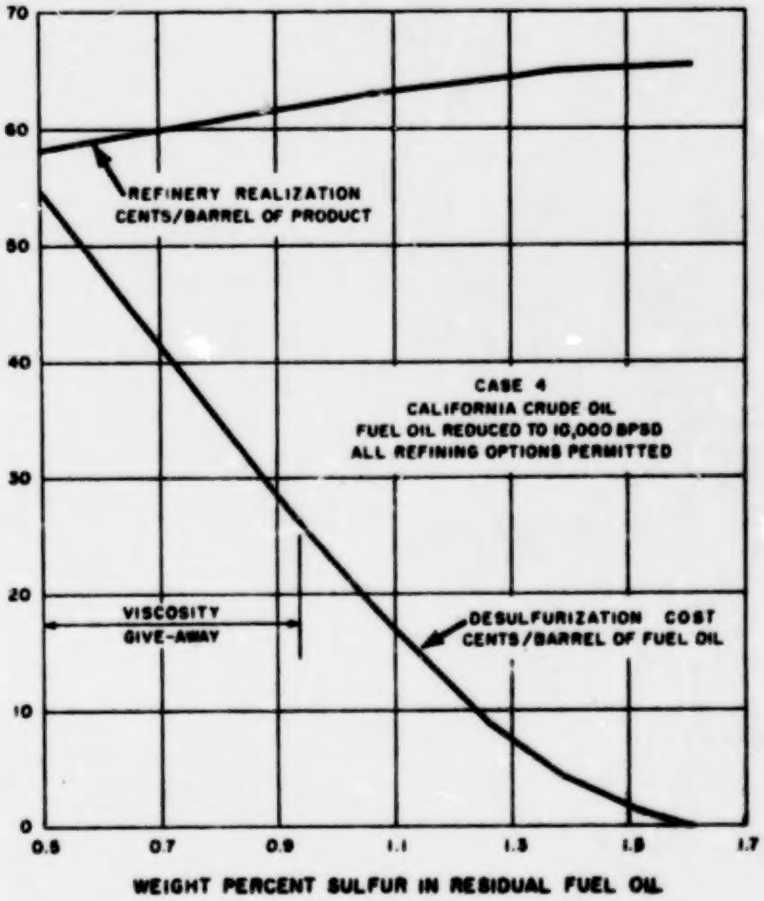
Case No. 1 is the base case in which the full range of refinery processes are available to deal with the sulfur problem. Desulfurization of the residual fuel oil was accomplished primarily by residual oil coking and hydrogen treating the cutter stocks. At low sulfur levels, the ratio of cutter stocks to residual oils was increased beyond the level required for viscosity blending alone.



Case Nos. 2 and 3 show the effect of excluding coking from consideration and thereby forcing the more expensive alternate of residual oil hydrodesulfurization to be used at all levels of sulfur in the residual fuel oil.

While the differential cost curve is slightly lower than in Case No. 1, the refinery realization curve is also lower. Case No. 3 is the same as Case No. 2 with the added restriction that hydrogen treatment of catalytic cracking feedstock was prohibited. No significant difference in cost resulted.





Case Nos. 2 and 3 serve to illustrate that coking and resid hydrodesulfurization are competitive processes for converting residual stocks to more valuable refinery products.

Case No. 4 repeats Case No. 1 except that the residual oil production is reduced from 24,000 BPSD to 10,000 BPSD. The cost of fuel oil desulfurization increases somewhat over Case No. 1, but there is also a striking improvement in refinery realization.

Case No. 5, which is not illustrated by graph, explores the profitability of fuel oil manufacture over a wide range of production levels while maintaining a constant 0.5 wt. percent sulfur specification. As the rate is cut from 32,478 BPSD in a series of steps to 10,000 BPSD, the realization rises from $0.279 per barrel of products to $0.531 per barrel. For the conditions of the study, the break-even price for incremental fuel oil production at 0.5 wt. percent is calculated to be $2.75. Below this price, there is economic pressure on the refiner to reduce production of this product.

Case No. 6 is a new base case for supplemental studies in which a constant 5,000 BPSD of Kuwait crude substitutes for part of the California crude. The mathematical data generated by the LP permitted separate costs of desulfurizing the high sulfur Kuwait residuum to be calculated. Figure 3 shows the resulting comparison between Kuwait and California crudes.

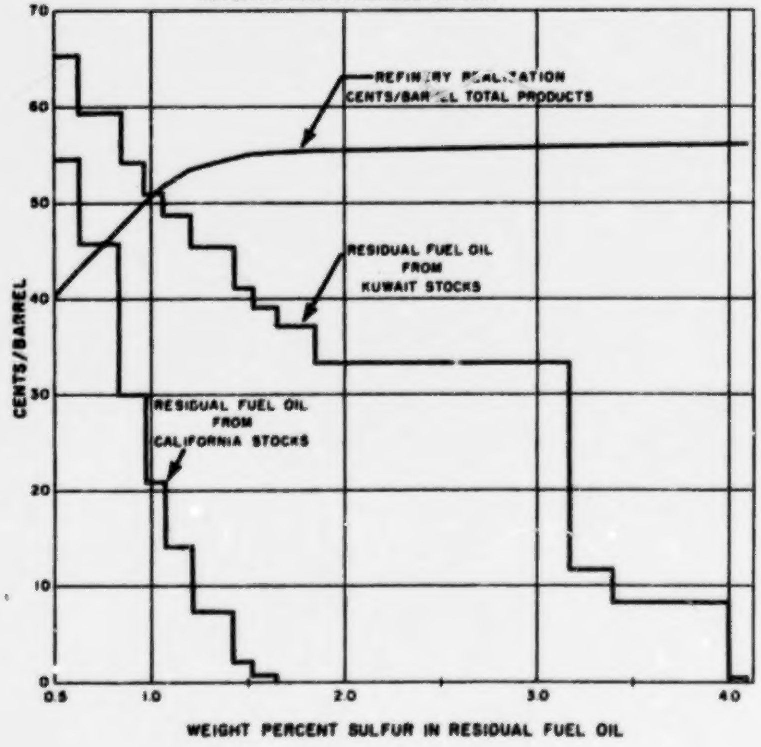
Also included in Case No. 6 was a small 500 BPSD increment of 4.5 percent sulfur "imported" fuel oil. The model refinery used this as a "crude" and the LP calculated a value or "shadow price" for it at various sulfur levels.

At the 0.5 wt. percent level, its value was $.80 lower than the price for desulfurized product fuel oil. This figure can be taken as a measure of the desulfurization cost.

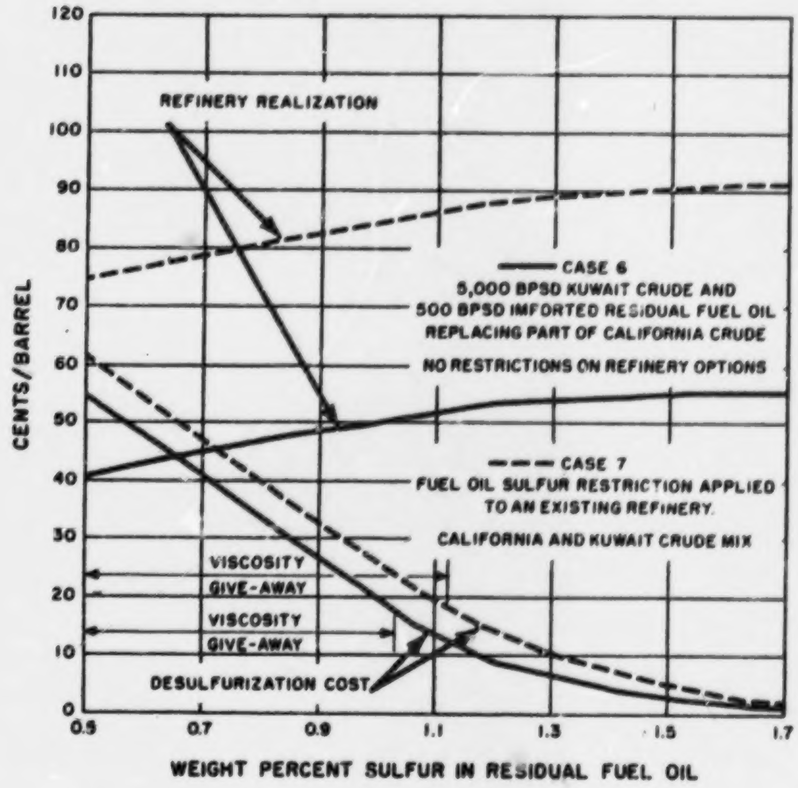
CASE 6
COMPARISON OF DESULFURIZATION COST FOR
CALIFORNIA AND KUWAIT RESIDUAL FUEL OILS

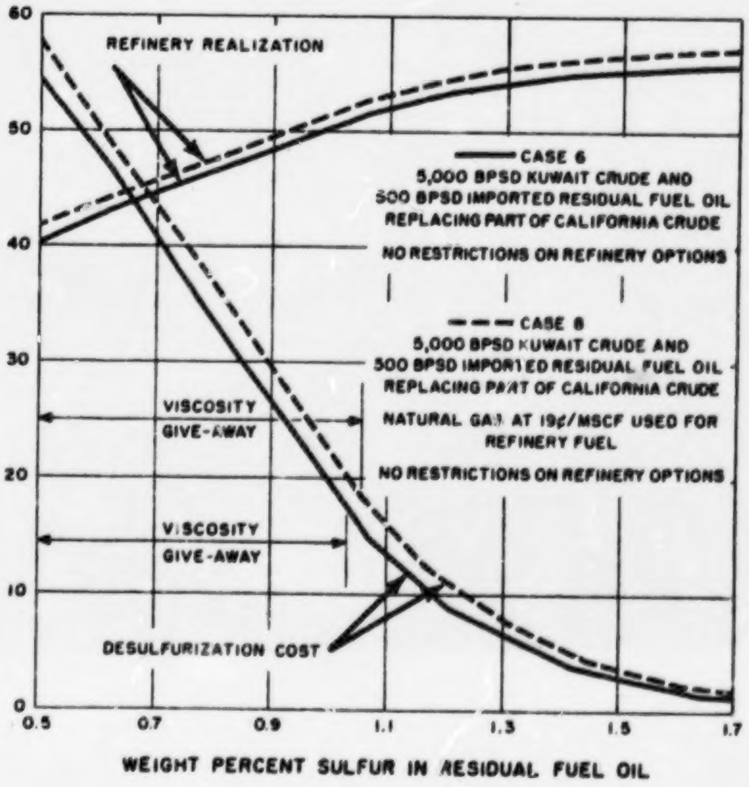
5000 BPSD KUWAIT CRUDE AND
500 BPSD IMPORTED RESIDUAL FUEL OIL
REPLACING PART OF CALIFORNIA CRUDE

NO RESTRICTIONS ON REFINERY OPTIONS



Case No. 7 considers the consequences of a restriction on fuel oil sulfur when it is applied to an existing refinery, optimized to process 100,000 BPSD of crude without limitation on fuel oil sulfur. It is assumed that the refinery is fully depreciated when the sulfur limit is applied. The incremental cost to lower the sulfur content of fuel oil is about 20 percent higher than for an all-new, optimized plant such as Case No. 6, but is not otherwise radically different. To conform to the 0.5 wt percent limitation, the refinery would have to add 10,000 BPSD of coking capacity and 14,000 BPSD of middle distillate hydrodesulfurization.





Case No. 8 shows the effect of making natural gas fuel available at low cost. When Case No. 8 is compared with Case No. 6 it is found that the availability of cheap natural gas improves the profitability of the refinery by reducing operating costs. The refinery configuration is not significantly affected by this change except that more hydrogen is used in processing.

It is of particular interest that the availability of cheap hydrogen did not reduce the cost of desulfurizing fuel oil. If anything, the relative attractiveness of alternate uses for the stocks blended into fuel oil were enhanced by the availability of cheap hydrogen with the result that fuel oil desulfurization cost was somewhat higher than in the case with higher priced hydrogen.

The overall conclusions of this study may be summarized as follows:

1. Manufacture of low sulfur residual fuel oil from high sulfur crudes requires an incentive pricing of $0.40 to $0.65 per barrel above fuel oil produced without sulfur restriction. This cost is increased about 20 percent if applied to an existing refinery.

2. Imported residual fuel oil can be regarded as crude and processed in a modern refinery. Its value is approximately $0.80 per barrel less than the desulfurized fuel oil product.

3. The availability of cheap natural gas reduces refining costs in general, but it does not appear to make fuel oil desulfurization more attractive.

THE NATURE OF THE PROBLEM

The nature and extent of the petroleum sulfur problem can be seen from Table I which is a tabulation of Bureau of Mines data in the several 1962 Mineral Industries Surveys published by the Bureau of Mines. The average sulfur content and API gravity of each of the products for each geographic area was considered to be representative of the total volume marketed in that district. Because of this and other assumptions in the calculations, no statistical accuracy can be claimed. However, the table does point up clearly the fact that the greater part of the sulfur released to the atmosphere comes from the burning of residual oils. The proportion is likely to show an increase in the near future, because competitive quality requirements are now forcing the refiners to desulfurize larger proportions of the other products. Further, the processes for desulfurizing these lighter stocks have been well studied and their place in the refining scheme is well established. For these reasons, this study has been confined entirely to residual oils.

A thorough literature survey by the Bureau of Mines indicates that hydrodesulfurization is the only refining process that appears promising for removing sulfur from residual petroleum oils. Several of the major oil companies are known to be developing residual oil desulfurization processes. However, only two commercial processes have been announced, the Gulf HDS process and the H-Oil process developed jointly by Hydrocarbon Research, Inc., and Cities Service Corporation.

It should be mentioned in passing that essentially complete removal of sulfur from residual petroleum oils by hydrogen treating is easily accomplished in the laboratory. However, these laboratory reactions require high hydrogen pressure, large ratios of catalyst to oil and large quantities of hydrogen. The commercial processes do not attempt complete desulfurization, use special catalysts and incorporate mechanical features which attempt to overcome these limitations. Even so, in the present state of development residual oil hydrodesulfurization is expensive in comparison with other refining processes.

Published figures indicate that the total costs for the desulfurization might be as high as $0.70 to $1.00 per barrel of residual oil processed. If it were assumed that this is an additive cost to be applied directly to the price of residual fuel oil, the process would be prohibitively expensive for many situations. However, an analysis which considers the desulfurization process by itself cannot give a true indication of the cost of the operation. The product oil from the hydrodesulfurizer actually consists of a wide range of materials no longer typical of the feed oil. From the refiner's viewpoint it can be regarded as a synthetic "crude" with a value determined by the products that can be prepared from it. Only a small fraction of the hydrodesulfurized product might be blended to fuel oil if the refinery were "optimized".

The inter-relationship between the various refinery processing units is extremely complex, and current practice is to approach these problems with a mathematical simulation or "model". This study has been largely directed toward creating and manipulating such a refinery model in which residual oil hydrodesulfurization is only one of several processes which can contribute to reducing sulfur content of residual fuel oil sulfur.

In approaching a study of this nature, attention must be directed to the material itself, the residual fuel oil. ASTM Grade No. 6 fuel oil is characterized generally by the absence of any specifications other than flash point (for safety in storage), viscosity (to insure pumpability), and the maximum content of water and sediments that can be tolerated in atomizing burners. The refinery thus has considerable latitude in the choice of components to blend in the product. In most cases the residual fuel oil sells for less than the cost of crude, and it is a marginal product kept to a minimum economic production level.

The degree of freedom that exists for the refinery and the nature of the material itself is best illustrated by some examples. For reasons to be explained later, a 23.8°API California crude was selected as a basis for much of this study. If this crude were processed in a simple topping refinery having only a crude unit and a catalytic reformer, the product distribution might be as follows:

Gasoline	14%
Lt. Distillates	8%
Residual Fuel Oil	76%
Operating Loss	2%
	100%

Sulfur content of the residual oil would be slightly in excess of 1.5 wt. percent.

In the above example, large quantities of middle distillates and gas oils must be retained in the residual fraction to meet the maximum viscosity specification. Application of modern refinery processing such as vacuum distillation and propane de-asphalting can reduce the heavy fraction to an asphalt residue amounting to only 14.4% on the original volume and at the same time recover oils that are valuable as feeds to other units. The asphalt that remains is a hard solid at room temperature. Unless it can be sold as asphalt, its disposal is a problem for the refiner. Only two options are generally available: the material may be blended and sold as fuel oil, or it may be converted to other products.

If it is to be sold as fuel oil, the refinery must blend the residual oils with sufficient "cutter stock" to meet the viscosity specification. Generally, this cutter stock is the lowest quality middle distillate available. On the assumption that it is a light catalytic cycle oil, approximately two volumes of cutter stock will be needed for each volume of the asphalt described above. The net refinery yield of residual fuel oil in this case is approximately 44%, the minimum that can be attained by blending alone without a residual oil conversion facility.

In nearly every instance where the refinery has facilities for fractionating the various products from the crude, the most economical fuel oil blends will consist of heavy pitch such as the propane precipitated asphalt described above and a low viscosity middle distillate. The fuel oil reduction that can be accomplished by selective blending is substantial and in the above case amounts to 32% on crude charged.

Because residual fuel oil usually sells for less than the cost of crude, the refiner is under pressure to reduce the quantity still further than is possible with these selective blending methods. Conversion processes include thermal coking, viscosity breaking and the residual oil hydrodesulfurization.

Of the three processes, only coking is capable of eliminating the fuel oil entirely. The oil is heated to approximately 850°F and allowed to flow slowly through a large coking drum where it decomposes to a hard carbonaceous residue and a wide boiling range "synthetic crude" which can be fractionated into various products. Coking, like most of the commercial non-catalytic cracking processes, has the disadvantage of producing low quality gasolines and unstable distillate fractions which must be further processed. Further, much of the sulfur in the residual oil is concentrated in the coke and limits its usefulness for metallurgical purposes. It is even marginal as a fuel since the anticipated high sulfur content would limit its applicability. Even with these limitations, however, coking is a widely used process, and may be economically attractive even when the coke product is not salable.

Viscosity breaking is a mild thermal cracking process which converts part of the residual oil to distillate products. In many instances, only the gasoline is removed, and the various gas oils and distillate stocks resulting from the cracking are retained in the fuel oil to act as "cutter stock". The net result is a significant reduction in the refinery yield of fuel oil because the amount of outside cutter oils is reduced. It is interesting to note that one Southwest refinery for many years followed the visbreaking operation with a high vacuum distillation step. This reduced the "black oil" to a hard pitch which was subsequently discarded in a canyon. Where the pitch can be disposed of, this viscosity breaking vacuum distillation scheme is an alternative to coking.

Residual oil hydrodesulfruization processes from the refiner's point of view resemble viscosity breaking and are sometimes referred to by the terms "hydro-visbreaking" or "resid hydrocracking". As descriptive names, these are in some respects superior to terms that suggest only sulfur removal. Regardless of the conditions employed or the type of process, the desulfurization reaction is accompanied by a substantial amount of cracking. Even simple removal of sulfur from a heavy molecule in a way that does not involve rupture of carbon-carbon bonds, reduces the molecular weight and increases volatility.

If these products of the hydrocracking and hydrodesulfurization reactions are retained in the fuel oil, the net effect is to reduce the overall refinery yield of residual fuel oil by reducing the need for outside cutter stock in a way analogous to viscosity breaking. Further, if the products from the hydrodesulfurization reaction are regarded as "synthetic crude" and fractionated to recover all of the gas oil and distillate fractions, it will be found that the volume of fuel oil can be reduced even further. The same selective blending procedure applied to virgin asphalt can be used with the residuum from the hydrodesulfurizer unit.

It is important to note, however, that the yield of desulfurized residuum from resid hydrodesulfurization processing is only a fraction of the original charge volume if all the gas oil and other distillable material is removed. Depending on the operating severity, catalyst activity, and crude type, the small amount of residuum remaining may still have a high sulfur content approaching that of the feedstock. This material may be blended to make a reduced quantity of fuel oil or it can in some cases be recycled to the hydrodesulfurization process so that there is no net production of fuel oil.

Residual oil hydrodesulfurization is a relatively new development, and only one plant is in operation. Therefore, nearly all the yield information in the literature consists of laboratory or pilot plant data. The yield data for this study came from two different sources. They serve to illustrate quite well the extremes in flexibility that exist for the design of a desulfurization unit.

The California residuum data were taken from the Galbreath and Johnson paper. The results are typical of a mild treatment of a refractory residuum. The sulfur content of the treated residuum (not the total treated product) rather closely matches the sulfur content of the original feedstock.

The Kuwait data were taken from the recent work of Beuther and Schmid. A very active catalyst was used, together with severe conditions. The result was a minimum yield of relatively low sulfur residuum.

Regardless of the type of process, the primary effect of residual oil hydrodesulfurization is a drastic reduction in volume of residual stock for fuel oil blending.

CRUDE SELECTION AND PRODUCT SLATE

In a study such as this, the selection of crude type and the product distribution is of fundamental importance. In the early phases of the work, some thought was given to setting up a model descriptive of the U. S. petroleum refining industry as a whole. Such a linear programming model is described by Manne in connection with a study on the capability of the industry to produce jet fuel. Unfortunately, Manne's model does not include any of the hydrogen treating processes that are essential to the current study. If these were included along with the necessary flexibility in processing operations that Manne's model did not need, the matrix would have several hundred equations, a simulation beyone the scope of this project.

Also, his model was based on the crudes from the first 25 of the major oil fields as listed in the Bureau of Mines survey for 1950. During the 12 years from that time to 1962, 50% of the fields had been dropped from the list to be replaced by many that had not been discovered at that time. Assays were not available for many of these crudes.

From a practical standpoint, residual oil hydrogen treating yield data are available only for a few crude types. Practically all of the work has been confined to Middle East, California or selected West Texas crudes, none of which are really typical of the average crude mix for the U. S. It is extremely important in an LP study to have a consistent set of premises for the simulation.

Because of the large amount of detail required for each crude type, the simulation had to be limited to one or two of these crudes. California crude was selected for the base case for two reasons. It was used in a recent study on fuel oil reduction methods by Slyngstad and Feigelman and some of their data could be adapted to this study. Also, the fuel oil made from California crudes will normally run close to the national average of 1.6 wt. percent sulfur. Results based on California crude should be fairly representative of what it will cost the industry as a whole for desulfurization of residual fuel oil.

An assay of a particular sample of Huntington Beach crude that closely corresponded to the average API gravity for total California production was available. Important properties are as follows:

Gravity, °API	23.8
Sulfur, Wt. Percent	1.3
Reid Vapor Pressure, PSI	1.1
Viscosity, SSU @ 100°F	149
Viscosity, SSU @ 80°F	246
Price, $/Barrel	2.46

The properties of the various fractions used in the LP may be found in the matrix coefficients as given in the appendix.

Product specifications and demands used for the matrix are given in Table 2. The product distribution corresponds to the approximate market breakdown in the West Coast for 1962 as presented by Slyngstad and Feigelman. These provide a realistic set of product demands corresponding to the crude gravity.

Since the primary purpose of the simulation was to study the effects of a sulfur specification restriction on residual fuel oil, the other products are considered to be "pools" with only average specifications typical for the material. The gasoline pool represents an average for premium and regular gasolines with an octane number slightly higher than the average being marketed today. The light distillate pool includes #1 burner oil, JP-5 jet fuel, kerosine, and light diesel. Heavy distillates include #2 burner oil and heavy diesel.

Specifications for these materials are limited only to boiling range and sulfur content. The sulfur contents are typical of what a refiner might normally set for products from this particular crude.

The LP matrix is purposely structured to eliminate any need for product prices in determining the cost of desulfurization. The matrix is set up to produce the required product slate for each case irrespective of prices, so that the only variables are the quantity and sulfur content of the fuel oil. In this way, the model is completely independent of the effects of product market prices. However, there is pressure on the refiner to reduce production of fuel oil, and this pressure can be measured in terms of gross realization relative to product prices. For this purpose the prices estimated by Slyngstad and Feigelman as refinery net-backs for the West Coast are used.

Two products from this model refinery which are not included in the above tabulation are coke and sulfur. Coke produced from this type of crude will have little value for most industrial purposes and because of its high sulfur content, it could not be burned as fuel in the marketing area served by the refinery. For this reason, the coke was given a negative value of $1.00 per ton to allow for a disposal expense.

Similarly, it was assumed that the refinery is not in the sulfur business, and no value was assigned to the byproduct sulfur. Assigning a value of, say, $20/ton would be reflected as a credit of $0.03 to $0.05 per barrel in favor of the resid hydrodesulfurization process in those areas where the sulfur could be marketed.

Cases 6, 7 and 8 expanded the scope of the study by replacing 5,000 BPSD of the California crude with Kuwait. Properties are as follows:

Gravity, °API	30.8
Sulfur, Wt. Percent	2.35
Reid Vapor Pressure, PSIG	8.1
Price, $/Barrel	2.00

Since the quantity of Kuwait crude was held constant in all cases, the price had no influence whatever on the LP matrix behavior. The only use made of the price of Kuwait crude was in calculating realization.

FORMULATING THE MATHEMATICAL MODEL

Inter-relationships between the refinery processing units are extremely complex, and until the recent development of mathematical programming methods, it was difficult to make meaningful comparisons between alternate refinery operations.

The most important and generally useful of the mathematical programming methods is Linear Programming (LP). It has become a standard technique for economic analysis in the petroleum industry. Once an LP matrix has been set up to describe a refinery complex, it can be programmed into a computer and used to calculate the economics of alternate modes of operating the refinery. It is possible to change product demands, specifications and refinery processing schemes with assurance that each solution will be accurately balanced and optimized. For this study, the Bonner and Moore LP code and GAMMA matrix generator were used on an IBM 7094 computer. The matrix generator has numerous features especially provided for oil refinery problems.

Formulation of the LP matrix is by far the most time consuming operation in a refinery simulation study. Considerable care must be taken to make sure that all reasonable alternate processing possibilities are provided. Also, it is extremely important that all of the data be internally consistent if valid results are to be obtained. For this study, the latter was taken as a guiding principle. The base case of the study was purposely narrowed to a single crude and marketing situation so that full attention could be given to constructing a matrix that would incorporate most of the process alternates that a refinery might elect to use. It was later expanded to evaluate another crude, and to permit handling some "imported" residual fuel oil.

In spite of its usefulness as a tool for economic analysis, linear programming does have limitations that influence the formulation of the problem and the interpretation of the results. The LP requires that a single cost figure be assigned to each possible operation, and this cost figure must be proportional to the "activity" of that operation. In blending operations, this requirement is accurately met; in others, it is approximated by restricting the range over which a cost applies.

The development of these "cost-per-barrel" processing costs is a matter of considerable judgment, and how they are formulated is closely related to the objectives of the study. Processing costs may be considered in three categories according to the simplifications that are made.

(1) General Overhead and Operating Labor. It is customary to ignore all costs of a general nature which cannot be identified with a particular product or operation. Since operating labor for a unit generally is practically independent of the feed rate, this is also omitted from consideration.

(2) Capital Related Costs. Where a unit already exists, it is usually assumed that amortization, interest, maintenance, taxes, and similar items are independent of the quantity of material being processed. These costs are excluded from the matrix. On the other hand, if new facilities must be built, it is proper to assign a cost penalty against each barrel of capacity that is needed.

(3) Utilities. Utilities, chemicals, catalyst, running royalties, and other factors that are proportional to the quantity of material processed are always included in the cost.

These distinctions are quite important, and the method of handling capital charges is often a source of confusion. In a study for a "grass roots" refinery where all new facilities are to be constructed, the capital related costs would be included in the matrix, and an optimum refinery configuration would result from the solution. After the refinery was constructed, a second LP solution might be made, with all of the capital cost factors excluded. This LP would find an optimum way of running the refinery which could be quite different from the original case in which individual unit sizes were determined. The reason for this is the change in objective. In the first case, the refiner has included cost factors to deliberately penalize the expenditure of capital. The objective is to build a refinery in which the expenses for capital write-off, crude and operation are balanced. Once the facilities have been constructed, there may very well be a more profitable way to run the refinery -- and the second LP solution will show this.

It is important to understand that the capital related charges, when included in the matrix, are really penalties to force the selection of low cost processing methods. They represent a balance of operating cost, cost of crude, and capital investment that is optimal for the refinery.

For the purpose of this study, two limiting extremes were set up. For most cases, each process unit considered for the LP matrix was penalized by a capital cost which included all maintenance, taxes and insurance, and "write-off" period of three years before taxes. The purpose of this was to optimize the refinery for minimum capital expenditure and at the same time establish an upper cost boundary that would reflect as severe a criterion as the industry might employ for incremental expenditure. As an alternate, the study also considers the case of a fully depreciated refinery which requires new facilities to reduce sulfur content of residual fuel product. In this case, the capital related charges were applied only to the new facilities.

The three year write-off period for establishing capital charges is normal from the industry's point of view. This is a handy rule of thumb, which corresponds to a true return of capital in about 5 years after taxes if the double declining balance method of depreciation is used. About 18 months will have elapsed from the time that a decision is made until new facilities could be constructed. The refinery accordingly is asked to commit capital for a period of 6-1/2 years. During this time, new technologies may be developed that obsolete the facilities, -- or the market conditions that call for the expenditure may change. If the tax laws are amended to permit faster depreciation of new facilities required for fuel oil desulfurization, refiners would find the investment more attractive.

In the present case where fuel oil desulfurization is being considered, the possibility of unexpected developments during that 6-1/2 year period is particularly great. Any nation-wide restriction that adds $0.45 or more per barrel to the cost of fuel oil is certain to cause a re-examination of energy sources by the industrial consumer, and it is entirely possible that the demand for residual fuel could decrease drastically. This would certainly prompt an investigation of alternate fuel sources. Such an investigation is beyond the scope of this report, but it would appear that both natural gas and coal would be possible alternates in existing fossilfueled heaters, and nuclear energy in new power plants. A preliminary inquiry has indicated that natural gas fuel for power plants will continue to be available for some time.

THE REFINERY FLOWSHEET

Figure 1 is a flowsheet showing the general flow arrangement for the refinery model. A considerable amount of time was spent attempting to prepare a fully descriptive flowsheet which could show all of the connecting streams between the refinery units. The resulting diagram was found to be unsuited to discussion purposes because of the almost unintelligible maze of lines. In fact, the LP matrix itself may be the most satisfactory means of representing the various flow paths in the model.

For discussion purposes, the simplified flowsheet, Figure 1, was prepared. The various refinery streams are grouped into five categories corresponding to the product pools plus one additional group for the refinery fuel gas and vapor pressure stocks.

Eleven main refinery processing units are shown, and these are grouped according to the general function. The groupings are self-explanatory, but it should be noted that the categories are to some extent arbitrary. In all of the hydrogen treating processes, some "cracking" takes place. Also, all of the cracking processes accomplish some degree of desulfurization.

The flow diagram is arranged specifically to emphasize an important fundamental principle of petroleum refining. Almost all of the cracking and reforming processes produce a "synthetic crude", consisting of some products in the same classification as the feed plus products in all of the lighter classifications. The catalytic reformer to which naphtha is fed produces a naphtha product and also byproduct gas and butanes classified as "light ends" on the diagram. Similarly, the delayed coker which charges the heaviest residual oils produces all of the lighter product classifications plus coke (which might be classified as a residual product).

Each of the units shown has wide latitude in accepting nearly all of the stocks in its feed classification that may be available. The solid lines depict most of the flow paths actually incorporated into the matrix. Several promising operations that were omitted for lack of data are shown with dotted lines. Treating propane deasphalted pitch by coking it, for example, might be quite attractive to a refinery where the respective units were available. Several other possibilities for processing the lighter products were not included because they are not important to this particular study. For example, refiners occasionally find it desirable to hydrogen treat and reform high sulfur naphthas from the catalytic cracking unit.

One particular process of considerable current interest is the Gas Oil Hydrodesulfurizer unit. Pre-treating the gas oil feed to the catalytic cracking unit is an effective alternate to treating the distillate products, and it has the added advantage that the gasoline yield and quality are improved at the same time.

The hydrocracker which converts heavier oils into gasoline and middle distillates or middle distillates to gasoline has become increasingly attractive to the modern refinery. A great deal of intensive development work is currently being done to lower costs and hydrogen consumption, as well as to increase the flexibility in accepting higher end point charge stocks. For this study, it was assumed that the maximum end point of the charge stock was 800°F. It was also assumed that the unit would be a typical two stage design with a "Hydro-denitrification" (HDN) pre-treatment section. The HDN section performs a severe hydrogen treating operation which removes both sulfur and nitrogen.

While only one "pool" is shown for the light ends, in most refineries there are two systems for handling these materials. The unsaturated propylene and butylenes resulting from coking and catalytic cracking are segregated so that they may be converted to gasoline by alkylation with isobutane or by polymerization. Saturated light ends may also be used for hydrogen plant feed if unsaturated hydrocarbons are excluded.

UTILITIES

Not shown on the flowsheet are the peripheral units and utilities which are incorporated into the matrix. An important question which is asked in connection with hydrogen consuming processes is the cost of hydrogen. It is frequently assumed that natural gas must be purchased for reforming into hydrogen, and on this basis the cost of desulfurization would depend on the cost of this raw material. As a complicating factor for this study, natural gas in most areas of the country competes with residual fuel oil for the bulk energy market. It is apparent that any restriction that increases the cost of fuel oil will increase the demand for natural gas, and this in turn will set a new and unpredictable price structure for that commodity.

To ascertain the availability of natural gas fuel for power generation and other industrial uses, inquiry was made to the Federal Power Commission who provided a copy of "Advisory Committee Report No. 21 on Fuels for Electric Generation" dated December 1963. On page 14 of this report, the following statement is made:

> "The supply of natural gas is not expected to impose any quantitative limitations on the use of natural gas for electric power generation between now and 1980. The use of natural gas for this purpose, however, can be expected to vary markedly among the several regions of the country. The extent to which natural gas is used for electric generation in each region will depend upon several factors, but primarily it will depend upon its price relative to the prices of alternative fuels. This, of course, will be affected by, among other things, the supply and demand relationship."

If the hydrogen unit is designed to do so, it can operate successfully on waste refinery gas. Also, various hydrogen containing vent streams from the various units can be collected and re-processed in the hydrogen plant with a considerable saving in overall fuel consumption. Based on these considerations, the refinery model was structured to be completely self-sufficient in terms of fuel, both as process energy and as hydrocarbon feed to the hydrogen unit.

Where large amounts of catalytic cracking and coking are employed, a refinery can approach a steam balance because these units produce byproduct steam. It was assumed that any refinery of the type depicted by this model would have a start-up stream generation facility independent of any size that might be calculated by a refinery balance. For this reason, no capital related charges were assessed against steam generation.

The matrix also includes a balance on electric power, and some simplification had to be made in this regard. It was assumed that power would be generated by an efficient "public utility" type installation, but that fuel for this purpose would be supplied by the refinery. The use of electricity was accordingly penalized at a constant 3 mil/Kwh rate for generation and distribution charges, but excluding the fuel value.

Considerable attention was given to structuring the problem so that it would operate in a closed environment. All fuel used for its own internal purposes for refinery fuel or electric power was assumed to meet the same sulfur specifications as that marketed. In other words, the refinery was required to burn a No. 6 fuel oil having the same specifications as imposed on the material for sale. This led to interesting behavior for those problems where the sulfur restriction was severe. At the high price developed for desulfurized fuel, the model found it profitable to burn iso-butane and propylene rather than convert them to gasoline as would normally be the case.

HYDROGEN BALANCE

The cost of hydrogen is the single largest item of expense connected with desulfurization processes. However, there is no simple direct relationship between the amount of hydrogen consumed and the resulting sulfur level in a specific product. The reason, of course, is that refinery products are blends of many components, and the hydrogen requirements for removing sulfur from the various blend components is variable. As has been pointed out, the residual fuel oil blends developed by the linear program largely consist of high sulfur residuum cut back with light distillate. Most of the sulfur reduction comes from treating the cutter stocks. As compared with direct resid hydrodesulfurization, this procedure results in substantial savings in hydrogen. While the resid HDS process requires in the order of 1000 SCF hydrogen per barrel of product, the requirements for desulfurization of cutter stock were generally in the range of 100 to 300 SCF per barrel.

The tabulated run data for the Case Studies (Tables Nos. 4 to 11) includes a figure for hydrogen plant capacity, and the differential hydrogen consumption figures can be easily computed for many of the cases. There is no particular pattern in the data, which simply indicated that hydrogen utilization was optimized along with other economic factors.

It must be stressed that the hydrogen plant sizes shown in the tabulations are only relative. Hydrogen cannot be stored and the hydrogen plant must have considerable reserve capacity for peaks. This reserve should be more or less constant for various situations and should not greatly influence the economics. The refinery matrix includes a complete hydrogen balance in which all by-product sources are utilized up to the limit of their availability. A practical refinery based on any of the processing schemes selected by the computer would undoubtedly require a substantially larger hydrogen plant than the balances indicate.

Hydrogen is a relatively expensive refinery raw material, and a considerable amount of effort can be justified in conserving its use. It is assumed in this problem that such steps would be taken. For example, the hydrogen consumption figures for the middle distillate hydrogen treater are taken from an actual design in which two stages of pressure let down are used so that the flash gas can be recovered and recycled.

Proper attention to design details can greatly improve the hydrogen utilization in a "hydrogen refinery". As Gwin points out, the purity requirements for the various processes does vary, and vent gases from one unit can serve as hydrogen feed to another. Even further, the various low pressure hydrogen rich flash gas streams that are available can be collected, compressed, and reprocessed through the hydrogen plant. The methane and ethane impurities are cracked to more hydrogen, while the hydrogen present simply is carried through unchanged. A substantial saving in fuel can be demonstrated by such a procedure.

Byproduct hydrogen from catalytic reforming is important as a source of hydrogen in a refinery. In this particular refinery model, the hydrogen from this source is relatively constant because the quantity of gasoline did not vary between cases. California crudes in general have high naphthene contents and give large byproduct hydrogen yields when the naphtha fractions are reformed. For other crudes, the overall size of the hydrogen plant would have to be increased, but this again would be a more or less constant factor.

The costs for hydrogen in this study assume a modern high pressure steam reforming unit, and are as low as can be reasonably expected at the present state of development. Because of the current interest in hydrogen processing in general, attention is being focused on hydrogen manufacturing process. However, it is doubtful that any significant "break-through" can occur which would greatly influence the results of this study. The reason is the large influence of fuel and other utility costs. Under the conditions of Case 1 to 7, the cost of hydrogen plant feed was equated to fuel produced by the refinery on the basis of its heating value. This resulted in a hydrogen cost of about $0.30 per MSCF, of which only $0.13 per MSCF is capital write-off. A large-scale, community type plant would, in effect, be able to reduce only this part of the cost.

RESULTS

The general conclusions have been given in the summary. In this section the results of each case study will be examined in greater detail.

THE CASE STUDIES

Case No. 1	24, 300 BPSD Fuel Oil Variable Sulfur Specification	Reference: Table No. 4

For this case, all of the products were held constant according to the demands and specifications in Table No. 4. Only the sulfur specification on No. 6 fuel oil was varied. In this way the effect on refinery costs attributable to the one single specification could be evaluated. The economic "optimum" sulfur content where there is no restriction is 1.63 wt. percent for this case. This was used as a base in computing the differential cost as the sulfur content is progressively reduced to a lower limit of 0.5 wt. percent. These differential costs are plotted in Figure 2.

The refinery configuration for this case is particularly interesting. Except for iteration No. 7-52 at 0.5 wt. percent sulfur, the Resid HDS unit was not used, and all desulfurization was obtained by a combination of coking, FCCU pre-treating (the Gas Oil HDS), and treatment of cutter stocks. Below approximately 0.95 wt. percent sulfur, the amount of low sulfur cutter stocks exceeds the viscosity blending requirements with the result that there is a give-away of viscosity. The viscosity of this fuel oil is lower than the specification requirement (45 SSF @ 122°F) and in some markets could command a small premium as a No. 5 fuel oil.

How the computer blended the final fuel oil product to arrive at 0.5% sulfur is interesting and shows typically how a refiner would deal with this problem.

Component	Wt. %S	Volume
Resid HDS byproduct middle dist.	0.5	106
Vacuum pitch	2.03	5,149
Resid HDS product pitch	1.7	163
Virgin heavy middle dist. (desulfurized)	0.12	10,289
FCCU Lt. Cycle oil (desulf. feed)	0.01	5,195
FCCU Hvy. Cycle oil (desulf. feed)	0.2	2,946
FCCU Decant oil (desulf. feed)	0.05	452
		24,300

The "Refinery Realization" as plotted in Figure 2 (and also in Figure 3 and 5) requires definition and qualification. The term as used in the context of this report has only relative significance because a number of costs facing an actual refinery operation have not been deducted. Typical of these are the capital charges associated with a number of off-site facilities such as start-up steam generation, crude and product storage, crude receiving and product shipment facilities and other costs, such as operating labor, factory overhead, laboratories and other controls, etc. These were omitted from consideration because they are affected only to a very minor extent if at all by the variants considered in the study.

Case Nos. 2 and 3 24,300 BPSD Fuel Oil Variable Sulfur Specification Reference: Tables 5 & 6

In Case No. 2, coking was excluded from the processing scheme. This forced the use of the resid hydrodesulfurization process as the only alternative to meet the increasingly severe restrictions on fuel oil sulfur. Comparison of Case No. 2 with Case No. 1 would then provide an assessment of the relative economics of coking and resid hydrodesulfurization.

In Case No. 3, both coking and pretreatment of fluid catalytic cracker feed were excluded. In this case, the only means available for middle distillate desulfurization would be the middle distillate hydrogen treater. Comparison with Case No. 2 would, therefore, provide an appraisal of the economics of catalytic cracker feed pretreatment versus middle distillate treatment.

Data for Case No. 2 are summarized in Table 5. It will be noted that the gross realization for Case 2 is uniformly lower than Case 1. By examination of the "Refinery Configuration" portions of Tables 4 and 5, it will be noted that the coking capacity called for in Case 1 has been replaced by resid hydrodesulfurization. The conclusion is that as far as California crude is concerned, coking is economically more attractive than resid hydrodesulfurization for processing California crude at any level of sulfur content. However, the incremental cost of producing low sulfur fuel in Case 2, as compared to its base case (Machine Iteration 10-73), is lower than in Case 1. (See also Figure 2)

Data for Case No. 3 are summarized in Table 6. If Case 3 is compared to Case 2, there is a rather small advantage in gross realization for Case 2, but the difference is very minor in the case of California crude. Again looking at the Refinery Configurations of Cases 2 and 3, the exclusion of Cat cracker feed pretreatment in Case 3 required a substantial increase in middle distillate hydrogen treatment capacity for Case 3. The slightly improved gross realization for Case 2 indicates that Cat cracker feed pretreatment is slightly more economical than middle distillate hydrogen treatment. The difference is very small and could reverse under other conditions. The incremental cost of fuel desulfurization of the Case 3 refinery compared to its base case (which is essentially the same as the Case 2 base case) is essentially the same as Case 2.

Case No. 4
10,000 BPSD Fuel Oil
Variable Sulfur Specification
All Processing Options Permitted
Reference:
Table No. 7

In Case No. 4, the product slate was altered from that used in Case Nos. 1, 2 and 3 by reduction of residual fuel oil production from 24,300 BPSD to 10,000 BPSD. All refining options available to Case No. 1 were permitted; thus, comparison of Cases 1 and 4 would provide an assessment of the profitability of manufacturing fuel oil.

Comparison of the data in Table 7 with those in Table 4 show that gross realization at all levels of sulfur in the fuel is higher for Case 4 than for Case 1. Fuel oil production is clearly unattractive, and there is substantial pressure to reduce production.

The incremental cost of desulfurizing residual fuel in the Case 4 refinery over its base case (Machine Iteration 8-66) is somewhat higher than for Case 1 (Machine Iteration 7-77). This is possibly related to the larger percentage of FCCU decant oil which is blended to fuel in the smaller volume of Case 4. Decant oil is a very low quality slop oil, high in sulfur, which usually has to be blended to fuel as the only practically available means of disposal.

Case No. 5
Constant Sulfur Specification of 0.5 wt. %
Variable Fuel Oil Quantity
All Processing Options Permitted
Reference:
Table No. 8

For this case, the residual oil sulfur specification was held at a constant value of 0.5 wt. percent while the total quantity of fuel oil was varied. Note that Iteration No. 8-52 of Case No. 4 represents a point at 10,000 barrels of residual fuel that can be considered a part of Case No. 5. Several points are worth noting: (1) As the production of fuel oil is increased, the amount of gas oil cracking needed decreased (FCCU and hydrocracker). Larger amounts of the feed stocks to these untis are consumed in the fuel oil blends. (2) At one point, the Resid Hydrodesulfurization unit entered the solution showing that it and coking are competitive processes, the choice of which will be governed by other factors than residual oil sulfur specification. (3) At the low residual oil yield levels, the hydrocracker becomes important as a means of using up distillate oils that otherwise would have gone to fuel oil blending. In all of the runs, the hydrocracker is important where middle distillates are in excess.

The significance of Case No. 5 lies in gross realization per barrel of refinery product figures. These vary greatly with the total fuel oil production.

Fuel Oil, BPSD	10,000	13,138	22,380	28,284	32,478
$ Realization	43,742	41,694	36,059	32,404	29,629
$ Loss/SD	--	2,048	7,683	11,338	14,113
$ Loss/Barrel	--	0.65	0.62	0.62	0.62

At the product values used for calculating the realization, each barrel of fuel oil produced above 10,000 BPSD represents a loss of gross realization of $0.60/barrel or more. The break-even point for 0.5 wt. percent sulfur No. 6 fuel oil is accordingly close to $2.75/barrel for the conditions of this study.

No computer runs were made to explore effect of fuel oil volume at high sulfur levels. A comparison between Case Nos. 1 and 4 shows, though, that at the 1.6 wt. percent sulfur level, there is a gain in refinery realization of about $2,300 in reducing fuel oil from 24,300 to 10,000 BPD. This corresponds to an incremental loss of realization of $0.16 for each barrel of fuel oil produced. Even without a sulfur restriction, reducing fuel oil appears attractive to the refiner.

Consideration of the above cases shows clearly what is already well recognized in the oil industry; production of No. 6 fuel oil is not a profitable business. It also explains why nearly one half of the No. 6 fuel oil consumed in the United States is imported.

Under the pricing used in this study ($2.14 per barrel for No. 6 fuel and $2.46 per barrel for crude), the refiner is under pressure to reduce fuel oil production without any restriction on sulfur content; he is, in effect, losing $0.16 for each barrel of No. 6 fuel he produces, if his production rate is about 25 percent of crude run. If a specification restricting fuel sulfur to 0.5 percent maximum were applied, the loss would rise to $0.60 as compared to alternate processing schemes which would reduce the production of No. 6 fuel. To prevent such a loss, a refiner would find strong justification for spending capital funds to purchase the process equipment required to reduce production of No. 6 fuel, and may even be able to justify going out of the No. 6 fuel business altogether.

It is also interesting to note that although the LP was given the option of propane deasphalting, this alternate was never selected.

Case No. 6 — Substitution of 5000 BPSD Kuwait Crude Oil in Crude Slate Plus Addition of 500 BPSD "Imported" Kuwait No. 6 Fuel Oil Variable Sulfur Specification

Reference: Table No. 9

In Case Nos. 6, 7 and 8, the effect of incorporating a second crude was studied. In addition, provision was made for including means of desulfurizing a relatively small amount of "imported" high sulfur No. 6 fuel oil made from Kuwait crude.

The LP matrix was also structured to provide in the products, an extra 1,000 BPSD of No. 6 fuel oil made entirely from Kuwait components. By following the shadow prices calculated by the LP program, the costs for refining the Kuwait crude, and desulfurizing the "imported" fuel oil could be determined on an incremental basis.

By holding the Kuwait components to incremental quantities, the basic structure of the refinery would not be so drastically changed as to make comparison with the non-Kuwait cases impractical.

The refinery models did take into account the special characteristics of Kuwait crude. Thus, special models were incorporated in the matrix for Kuwait crude distillation, catalytic reforming including the pretreat, catalytic cracking, residual oil hydrodesulfurization and coking. To prevent excessive complication of the LP matrix, the Kuwait models were simplified by limiting the reformer to only one level of octane improvement, and the cat cracker to one conversion level. These simplifications were possible because with the relatively small amount of Kuwait crude in the feed, the approximation would introduce a negligible error; the advantage of the simplification was an appreciable reduction in the machine time for reaching a solution.

The product slate for Case No. 6 (and also for Case Nos. 7 and 8) was as follows:

	BPSD
Motor Gasoline	40,300
Light Middle Distillate	8,000
Heavy Middle Distillate	16,900
Residual Fuel Oil	24,300
Special Kuwait Residual Fuel Oil	1,000
	90,500

The pertinent results of Case No. 6 are summarized in Table No. 9, and in Figures 3, 4 and 5.

The gross realization per barrel of product given in Table No. 9 for Case No. 6 is generally comparable to Case No. 1 - they differ only by about $0.01 per barrel. The incremental cost of fuel oil desulfurization is, as might be expected, higher than in Case 1, reflecting the higher quantity of sulfur in the feed.

Of much greater interest are the data plotted in Figures 4 and 5. In Figure 4 are given the "Shadow Prices" for Kuwait fuel oil, California fuel oil, and the "imported" high sulfur fuel oil, together with the shadow cost of removing 1 percent sulfur from one barrel of California and Kuwait crude. These data are all plotted as a function of the percentage of sulfur and each of these fuels.

Referring first to the shadow prices of Kuwait, California and "imported" fuel oil, the following conclusions may be drawn:

1. Kuwait fuel oil is relatively expensive material to refine to low sulfur content. This is apparently due to the increased requirement for resid hydrodesulfurization and middle distillate hydrogen treatment as compared to California crude. Interestingly, the shadow prices of Kuwait and California crude become equal at 0.5 percent sulfur, indicating that at this sulfur level, any incremental production could be made interchangeably from either crude, as far as cost is concerned, where the production rate is 25,300 BPSD of total fuel.

2. **Imported high sulfur fuel oil is an undesirable feed. At 0.5% sulfur specification for fuel oil, the shadow price of this material is $2.04, compared to an assumed market price of $2.14 and a shadow cost of manufacturing specification fuel of $2.84 per barrel.**

The data in Figure 5 were obtained by integrating the shadow prices given in Figure 4 over the whole range of sulfur reduction. Figure 5, therefore, gives the total cost of desulfurization of Kuwait and California residual fuel, from their natural sulfur contents down to 0.5 percent sulfur. Thus, Kuwait fuel would cost an incremental $0.658 per barrel for desulfurization to 0.5 percent compared to $0.54 for the balance of the fuel. This has been labelled "California" in Figures 4 and 5 - actually it includes Kuwait components not accommodated in the 1,000 BPSD of special Kuwait fuel. The difference in incremental desulfurization cost for "California" crude in Case 6 from that of Case 1 is partly due to this contamination with Kuwait.

The shadow cost of desulfurizing residual fuel given in the lower portion of Figure 4 requires further comment. It will be noted that the shadow cost for removal of 1% sulfur from a barrel of fuel oil is higher for California crude over the range from about 1.3% to 0.5%. In interpreting these curves, it must be remembered that the shadow cost is a derivative or slope of the total cost at some point. In the above range, the refiner would be installing the bulk of his desulfurization facilities for California crude, whereas he would have to install them much earlier for Kuwait.

Case No. 7 Existing Refinery Operation Under Progressively Lowered Maximum Limits on Sulfur Content of No. 6 Fuel

Reference: Table No. 10

Case 7 simulates a modern 100,000 barrel refinery built with no regard for residual fuel oil sulfur and develops the various processing capacities the owner of such a refinery would have to install in order to comply with progressively more restrictive limits on the sulfur content of No. 6 fuel oil.

The matrix includes Kuwait crude and Kuwait fuel in the feed as in Case No. 6 with the additional feature that a base capacity was established for each of the major refinery processing units. This basic refinery was set up after examination of several preliminary computer runs. It is a computer optimized refinery and is typical of the type of modern refinery that might be built for the product mix and crude slate used in this study. The fluid catalytic cracking unit, the catalytic reformer, and the naphtha hydrotreater units were purposely set at figures somewhat larger than the optimum called for by the computer solution. This is also true of the crude unit which was set at a nominal capacity of 100,000 barrels per day. The adjustments in all cases were relatively small and intended to give the refinery the operating flexibility that the refinery designers would specify for a real situation. Furthermore, all units in this basic refinery are assumed to be fully depreciated so that no capital charges apply. An LP matrix was then structured to permit the construction of new facilities as required to meet increasingly severe sulfur specifications in the residual fuel. For these new facilities a capital investment charge was included.

The original refinery showed a satisfactory hydrogen balance sufficient for the modest size hydrocracker and middle distillate hydrotreater. By-product hydrogen from the reformer was adequate for these purposes. The matrix was given the option to build hydrogen manufacturing facilities as needed. Results of Case No. 7 are not radically different from those obtained in the other cases presented. The cost of sulfur removal as shown in Figure 3 is about 10-20% higher than for the new refinery cases. This indicates that the position of the oil company having an existing refinery is not radically different from the "grass roots" cases previously studied. Figure 6 summarizes the results of this particular study by showing the incremental additions to each unit plotted against the weight percent fuel oil sulfur specification. Further information is given in Table No. 10. These results are entirely consistent with the others that have been presented. As can be seen, delayed coking and middle distillate hydrogen treating dominate. At 0.5 wt. percent sulfur specification the refinery must add a 10,000 barrel delayed coker and almost 14,000 barrels of middle distillate hydrogen treating. A nominal amount of new catalytic cracking is required to balance gasoline production. This does not appear to be particularly significant and simply indicates that possibly the base refinery may not have had sufficient cracking capacity for adequate flexibility.

As has been explained in Cases 2 & 3, residual oil hydrotreating and coking are competitive processes. Which one is most economical depends on the particular circumstances of crude type, hydrogen cost and the importance of capital expense factors. Both processes can be used to effect either a reduction in total fuel oil production or to reduce its sulfur level. This is illustrated nicely by the different behavior of California and Kuwait residual stocks. Coking was preferred in almost all cases with California stocks, while residual oil desulfurization was found to be most economical for Kuwait stocks. There are two basic reasons for this:

1. When the residual hydrodesulfurization process is carried out on Kuwait residuum, desulfurization reactions assume a much more important position than in the case of California residuum, probably because of the large amount of sulfur present in Kuwait. Each cubic foot of hydrogen consumed removes a greater weight of sulfur from Kuwait than from California stock.

2. The coker distillates from undesulfurized Kuwait residuum are much higher in sulfur than comparable materials from California, and require more severe processing to convert into salable products.

Taken together, these two factors favor residuum hydrodesulfurization for Kuwait stocks.

With only one exception, there is no indication that the type of crude would have any significant effect on the way in which units would have to be added to an existing refinery such as is considered in Case 7. This single exception is the hydrogen plant. The availability of highly naphthenic naphthas from California crude produces relatively large yields of by-product hydrogen from the reformer. Thus, with a more paraffinic crude input, hydrogen plant sizes would be larger than indicated here.

Case No. 8 — Cheap Natural Gas Available — Reference: Table No. 11

To cover the situation where cheap natural gas fuel is available, and hence, the possibility of cheap manufactured hydrogen, the refinery of Case No. 6 was re-optimized on the basis that natural gas would be available at a cost of $0.19 per thousand cubic feet.

There is no indication that the availability of low cost hydrogen makes it any more attractive for the refinery to desulfurize No. 6 fuel oil. In fact, if Case Nos. 6 and 8 are compared, it is evident that the principal effect of cheap natural gas is to improve the refinery profitability by reducing operating costs, mostly by reducing crude intake.

The pertinent data are summarized in Table 11 and Figure 3.

Examination of these data, and comparing them with Case No. 6 leads to the following conclusions:

1. Availability of cheap natural gas does not make any significant change in the optimum size of a coker.

2. Hydrocracking is substantially more favorable when low cost hydrogen is available, and tends to displace an equal quantity of fluid catalytic cracking. This applies even though there has been no change in the product slate.

3. Low cost hydrogen does not make a significant difference in the optimum sizing of resid hydrodesulfurization. A more drastic reduction in hydrogen cost might tip the balance in favor of resid hydrodesulfurization over coking however.

TABLES

TABLE NO. 1

DISPOSITION OF SULFUR IN REFINED PRODUCTS

YEAR OF 1962

	Product - 1000 BPD						Sulfur		
	Refined in Area	Imported	Exported	Net Consumed	#/Bbl	Tons/Day	%Content	Tons/Day	Approx. % Total Sulfur Burned
Composite Average - Total U. S. (Excluding Rocky Mountain Region)									
Gasoline	4,146	38	- 18	4,166	254	529,083	.043	228	5.0
Kerosine (Incl. Comm. Jet)	422	18	- 1	439	283	62,119	.079	49	1.1
Military Jet Fuel	261	30	---	291	275	40,013	.067	27	0.6
Distillate Fuel Oil	1,900	32	- 23	1,909	294.5	281,100	.213	599	13.2
Residual Fuel Oil	769	722	- 35	1,456	348.7	253,854	1.428	3,625	80.1
Asphalt	281	18	- 2	297					
All Other	787	95	- 84	798					
Totals	8,566	953	-163	9,356		1,166,169		4,528	100.0

TABLE 2

PRODUCT SPECIFICATIONS USED IN LP MODEL

Motor Gasoline - Total Pool	
Research Octane (F-1)	95.0
Reid Vapor Pressure, PSI	10.0
Light Middle Distillate	
Max. Wt. % Sulfur	0.25
Minimum B. P., °F	330
Max. B. P., °F	540
No Coker of FCCU Oils Permitted	
Heavy Middle Distillate	
Max. Wt. % Sulfur	0.50
Minimum B. P., °F	330
Max. B. P., °F (Cracked)	650
Max. B. P., °F (Virgin)	700
Residual Fuel Oil	
Max. Wt. % Sulfur	0.5-1.7
Max. Viscosity, SSF at 122°F	175
Minimum I. B. P., of Blend Stock °F	330

PRODUCT DEMANDS

Based on approximately 100,000 BPD of Crude

	BPSD	(Note 2) Value $/Barrel
Motor Gasoline	40,300	$ 4.95
Light Middle Distillate	8,000	$ 3.02
Heavy Middle Distillate	16,900	$ 3.02
Residual Fuel Oil (Note 1)	24,300	$ 2.14

Note 1 - Quantity of residual fuel oil is varied in the same LP runs. Other products are always kept constant.

Note 2 - Estimated values in refinery storage.

TABLE NO. 3

ABRIDGED SPECIFICATIONS FOR FUEL OILS

GRADE NO.	1	2	4	5	6
Test					
Flash Pt., °F Min.	100	100	130	130	150
Pour Pt., °F Max.	-	20	20	-	-
Water & Sediment, % Max.	tr.	0.10	0.50	1.00	2.00
Carbon Resid, 10% BTMS, % Max.	0.15	0.35	-	-	-
Ash, Wt. %, Max.	-	-	0.10	0.10	-
Distillation					
10% Max.	420	-	-	-	-
90% Max.	550	640	-	-	-
90% Min.	-	540	-	-	-
Viscosity					
Kinematic, CS, 100F, Max.	2.2	(3.6)	(26.4)		
Kinematic, CS, 100F, Min.	1.4	(2.0)	(5.8)	(32.1)	
SUS, Sec., 100F, Max.		37.93	125		
SUS, Sec., 100F, Min.		32.6	45	150	
SFS, Sec., 122F, Max.				40	300
SFS, Sec., 122F, Min.					45

Source: ASTM Tentative Standard D-396-61T

Note: Numbers in parenthesis are equivalent kinematic viscosities.

CASE NO. 1 EFFECT OF VARYING SULFUR CONTENT OF RESIDUAL FUEL OIL TABLE NO. 4

Machine Iteration Number	7-52	7-54	7-58	7-66	7-71	7-77
Wt. % Sulfur in Fuel Oil	0.5	1.04	1.12	1.26	1.40	1.63
Value of Products - $/SD	326,685	326,685	326,685	326,685	326,685	326,685
Less: Cost of Crude	239,351	235,835	235,768	235,206	235,011	235,582
Capital Related Costs	39,201	35,788	34,625	33,993	33,503	32,764
Variable Operating Costs	13,107	11,906	11,612	11,744	11,793	11,414
Gross Realization - $/SD	35,026	43,156	44,680	45,742	46,378	46,925
Crude Run, BPSD	97,297	95,868	95,841	95,612	95,533	95,765
Product Slate, BPSD	89,500	89,500	89,500	89,500	89,500	89,500
Realization/Crude Run - $/B	0.360	0.450	0.466	0.478	0.485	0.490
Realization/Total Products $/B	0.391	0.482	0.499	0.511	0.518	0.524
Crude Cost/Bbl Product	2.674	2.635	2.634	2.628	2.626	2.632
Capital & Variable Cost/Bbl Product	0.584	0.533	0.517	0.511	0.506	0.494
Δ Crude Cost/Bbl Product	+ 0.042	+ 0.003	+ 0.002	- 0.004	- 0.006	Base
Δ Capital & Variable Cost/Bbl Product	+ 0.090	+ 0.039	+ 0.023	+ 0.017	+ 0.012	Base
Total Δ Cost/Bbl Product	+ 0.132	+ 0.042	+ 0.025	+ 0.013	+ 0.006	Base
Total Δ Cost/Bbl Fuel Oil	+ 0.486	+ 0.155	+ 0.092	+ 0.048	+ 0.022	Base
REFINERY CONFIGURATION						
Crude Unit, BPSD	97,297	95,868	95,841	95,612	95,533	95,765
Reformer Pretreater, BPSD	22,848	21,062	20,318	20,221	20,217	19,564
Reformer, BPSD	20,097	19,345	19,626	19,623	19,562	19,932
FCCU Feed Pretreater, BPSD	29,886	18,524	7,116	7,673	8,850	0
Fluid Catalytic Cracking Unit, BPSD	22,613	22,925	21,028	21,278	22,052	19,793
Middle Distillate Hydrogen Treater, BPSD	16,865	16,569	19,969	16,703	12,820	11,923
Delayed Coking Unit, BPSD	23,712	17,824	17,846	17,948	18,233	19,605
Alkylation Unit, BPSD	2,981	2,993	2,839	2,986	3,084	3,003
Resid. HDS Unit, BPSD	815	0	0	0	0	0
Hydrocracking Unit, BPSD	1,007	2,679	3,981	4,010	3,891	4,960
Hydrogen Plant, MMSCF/SD	9.203	4.424	0	0	0	0
Sulfur Plant, M LB/SD	261	210	183	181	189	142

CASE NO. 2 **EFFECT OF VARYING SULFUR CONTENT OF RESIDUAL FUEL OIL** **TABLE NO. 5**

Delayed Coking Excluded

Machine Iteration Number	10-51	10-56	10-59	10-62	10-68	10-72	10-73
Wt. % Sulfur in Fuel Oil	0.5	0.639	0.777	1.022	1.135	1.258	1.276
Value of Products - $/SD	326,685	326,685	326,685	326,685	326,685	326,685	326,685
Less: Cost of Crude	230,310	227,506	227,302	226,974	226,846	226,719	226,377
Capital Related Costs	51,767	50,087	48,285	46,433	46,003	46,031	46,248
Variable Operating Costs	13,282	12,738	12,413	12,306	12,245	12,174	12,298
Gross Realization - $/SD	31,326	36,354	38,685	40,972	41,591	41,761	41,762
Crude Run, BPSD	93,622	92,482	92,399	92,266	92,214	92,162	92,023
Product Slate, BPSD	89,500	89,500	89,500	89,500	89,500	89,500	89,500
Realization/Crude Run - $/B	0.335	0.393	0.419	0.444	0.451	0.453	0.454
Realization/Total Products - $/B	0.350	0.406	0.432	0.458	0.465	0.467	0.467
Crude Cost/Bbl Product	2.573	2.542	2.540	2.536	2.535	2.533	2.529
Capital & Variable Cost/Bbl Product	0.727	0.702	0.678	0.656	0.651	0.650	0.654
Δ Crude Cost/Bbl Product	+ 0.044	+ 0.013	+ 0.011	+ 0.007	+ 0.006	+ 0.004	Base
Δ Capital & Variable Cost/Bbl Product	+ 0.073	+ 0.048	+ 0.024	+ 0.002	- 0.003	- 0.004	Base
Total ΔCost/Bbl Product	+ 0.117	+ 0.061	+ 0.035	+ 0.009	+ 0.003	+ 0.0	Base
Total ΔCost/Bbl Fuel Oil	+ 0.431	+ 0.225	+ 0.129	+ 0.033	+ 0.011	+ 0.0	Base
REFINERY CONFIGURATION							
Crude Unit, BPSD	93,622	92,482	92,399	92,266	92,214	92,162	92,023
Reformer Pretreater, BPSD	20,778	20,420	19,977	19,470	19,292	19,242	19,343
Reformer, BPSD	17,511	17,173	16,793	16,717	16,710	16,739	16,482
FCCU Feed Pretreater, BPSD	5,323	667	0	0	0	0	0
Fluid Catalytic Cracking Unit, BPSD	28,335	29,217	29,373	29,082	28,864	29,647	31,637
Middle Distillate Hydrogen Treater, BPSD	22,912	23,805	22,847	18,405	15,656	12,229	10,951
Alkylation Unit, BPSD	2,079	2,041	2,182	2,309	2,336	2,320	2,445
Resid. HDS Unit, BPSD	28,555	27,179	24,996	23,282	23,237	23,947	25,110
Hydrocracking Unit, BPSD	0	0	387	1,041	1,276	1,351	726
Hydrogen Plant, MMSCF/SD	20.848	16,923	15,243	13,671	13.277	13.414	13.313
Sulfur Plant, M LB/SD	233	217	207	190	182	185	190

CASE NO. 3 EFFECT OF VARYING SULFUR CONTENT OF RESIDUAL FUEL OIL TABLE NO. 6

Delayed Coking and FCCU Pretreating Excluded

Machine Iteration Number	12-50	12-55	12-58	12-60	
Wt. % Sulfur in Fuel Oil	0.5	0.660	0.794	1.022	At higher Sulfur levels, results are identical to Case 2
Value of Products - $/SD	326,685	326,685	326,685	326,685	
Less: Cost of Crude	229,882	227,489	227,274	226,974	
Capital Related Costs	52,491	49,770	48,073	46,433	
Variable Operating Costs	13,408	12,647	12,382	12,306	
Gross Realization - $/SD	30,904	36,779	38,956	40,972	
Crude Run, BPSD	93,448	92,475	92,388	92,266	
Product Slate, BPSD	89,500	89,500	89,500	89,500	
Realization/Crude Run - $/B	0.331	0.398	0.422	0.444	
Realization/Total Products $/B	0.345	0.411	0.435	0.458	
Crude Cost/Bbl Product	2.568	2.542	2.539	2.536	
Capital & Variable Cost/Bbl Product	0.736	0.697	0.676	0.656	
Δ Crude Cost/Bbl Product	+ 0.039	+ 0.013	+ 0.010	+ 0.007	
Δ Capital & Variable Cost/Bbl Product	+ 0.082	+ 0.043	+ 0.022	+ 0.002	
Total ΔCost/Bbl Product	+ 0.121	+ 0.056	+ 0.032	+ 0.009	
Total ΔCost/Bbl Fuel Oil *	+ 0.446	+ 0.206	+ 0.118	+ 0.033*	
REFINERY CONFIGURATION					
Crude Unit, BPSD	93,448	92,475	92,388	92,266	
Reformer Pretreater, BPSD	19,623	20,310	19,931	19,470	
Reformer, BPSD	16,971	16,920	16,782	16,717	
Fluid Catalytic Cracking Unit, BPSD	30,141	29,334	29,369	29,082	
Middle Distillate Hydrogen Treater, BPSD	31,049	23,984	22,686	18,405	
Alkylation Unit, BPSD	2,077	2,069	2,197	2,309	
Resid. HDS Unit, BPSD	28,502	26,869	24,729	23,282	
Hydrocracking Unit, BPSD	0	0	445	1,041	
Hydrogen Plant, MMSCF/SD	18.189	16.456	15.069	13.671	
Sulfur Plant, M LB/SD	226	214	206	190	

* Base is machine iteration No. 10-73 of Case No. 2

CASE NO. 4 **EFFECT OF VARYING SULFUR CONTENT OF RESIDUAL FUEL OIL** **TABLE NO. 7**

Constant Product Slate, 10,000 BPSD Residual Fuel Oil

Machine Iteration Number	8-52	8-53	8-57	8-58	8-60	8-62	8-66
Wt. % Sulfur in Fuel Oil	0.5	0.939	1.069	1.262	1.384	1.512	1.610
Value of Products - $/SD	296,083	296,083	296,083	296,083	296,083	296,083	296,083
Less: Cost of Crude	202,559	201,181	201,174	201,201	201,031	201,036	201,075
Capital Related Costs	37,079	35,955	35,398	34,565	34,244	33,930	33,794
Variable Operating Costs	12,703	12,305	12,168	11,961	12,012	12,030	11,989
Gross Realization - $/SD	43,742	46,642	47,343	48,356	48,796	49,087	49,225
Crude Run, BPSD	82,341	81,781	81,778	81,789	81,720	81,722	81,738
Product Slate, BPSD	75,200	75,200	75,200	75,200	75,200	75,200	75,200
Realization/Crude Run - $/B	0.531	0.570	0.579	0.591	0.597	0.601	0.602
Realization/Total Products $/B	0.582	0.620	0.630	0.643	0.649	0.653	0.655
Crude Cost/Bbl Product	2.693	2.675	2.675	2.675	2.673	2.673	2.674
Capital & Variable Cost/Bbl Product	0.662	0.642	0.632	0.619	0.615	0.611	0.608
Δ Crude Cost/Bbl Product	+ 0.019	+ 0.001	+ 0.001	+ 0.001	- 0.001	- 0.001	Base
Δ Capital & Variable Cost/Bbl Product	+ 0.054	+ 0.034	+ 0.024	+ 0.011	+ 0.007	+ 0.003	Base
Total ΔCost/Bbl Product	+ 0.073	+ 0.035	+ 0.025	+ 0.012	+ 0.006	+ 0.002	Base
Total ΔCost/Bbl Fuel Oil	+ 0.549	+ 0.263	+ 0.188	+ 0.090	+ 0.045	+ 0.015	Base
REFINERY CONFIGURATION							
Crude Unit, BPSD	82,341	81,781	81,778	81,789	81,720	81,722	81,738
Reformer Pretreater, BPSD	19,321	18,428	18,064	17,518	17,465	17,484	17,445
Reformer, BPSD	19,062	18,686	18,829	19,063	18,926	18,957	18,990
FCCU Feed Pretreater, BPSD	19,797	14,326	8,730	0	0	0	0
Fluid Catalytic Cracking Unit, BPSD	21,985	22,258	21,316	19,836	21,461	21,197	21,029
Middle Distillate Hydrogen Treater, BPSD	12,191	12,298	13,964	16,762	15,466	14,397	13,273
Delayed Coking Unit, BPSD	22,887	20,570	20,629	20,769	20,822	21,237	21,401
Alkylation Unit, BPSD	3,423	3,435	3,360	3,239	3,317	3,384	3,388
Hydrocracking Unit, BPSD	4,507	5,279	5,920	6,904	6,734	6,734	6,824
Hydrogen Plant, MMSCF/SD	8.607	6.781	4.613	1.248	0.741	0.531	0.343
Sulfur Plant, M LB/SD	236	217	204	184	186	183	179

CASE NO. 5

TABLE NO. 8

EFFECT OF VARYING RESIDUAL FUEL OIL PRODUCTION AT CONSTANT SULFUR CONTENT

Varying Fuel Oil Production at 0.5 Wt. % Sulfur

Machine Iteration Number	8-52	9-73	9-74	9-75	9-76
Heavy Fuel Oil - BPSD	10,000	13,138	22,380	28,264	32,478
Value of Products - $/SD	296,083	302,798	322,576	335,168	344,186
Less: Cost of Crude	202,559	210,701	234,698	248,956	261,631
Capital Related Costs	37,079	37,484	38,640	40,471	39,494
Variable Operating Costs	12,703	12,919	13,179	13,337	13,432
Gross Realization - $/SD	43,742	41,694	36,059	32,404	29,629
Crude Run - BPSD	82,341	85,651	95,406	101,202	106,354
Product Slate - BPSD	75,200	78,338	87,580	93,464	97,678
Realization/Crude Run - $/B	0.531	0.487	0.378	0.320	0.279
Realization/Total Products $/B	0.582	0.532	0.412	0.347	0.303
REFINERY CONFIGURATION					
Crude Unit, BPSD	82,341	85,651	95,406	101,202	106,354
Reformer Pretreater, BPSD	19,321	20,270	22,456	23,657	24,558
Reformer, BPSD	19,062	19,360	20,048	20,203	21,217
FCCU Feed Pretreater, BPSD	19,797	22,964	29,370	30,953	28,749
Fluid Catalytic Cracking Unit, BPSD	21,985	22,017	22,341	23,176	20,920
Middle Distillate Hydrogen Treater, BPSD	12,191	13,073	16,040	18,566	22,290
Delayed Coker, BPSD	22,887	23,187	24,182	22,943	26,599
Alkylation Unit, BPSD	3,423	3,319	3,063	2,813	2,650
Resid. HDS Unit, BPSD	0	0	0	2,497	0
Hydrocracking Unit, BPSD	4,507	3,617	1,495	0	0
Hydrogen Plant, MMSCF/SD	8.607	8.824	8.915	9.798	6.015
Sulfur Plant, M LB/SD	236	242	259	267	264

CASE NO. 6

EFFECT OF INCLUDING 5000 BPSD OF KUWAIT CRUDE AND 500 BPSD OF KUWAIT NO. 6 FUEL OIL IN REFINERY FEED

TABLE NO. 9

Machine Iteration No.	18-67	18-71	18-73	18-75	18-77	18-85	18-89	18-94	18-98	18-104	18-105	18-106	18-108
Wt. % Sulfur in Fuel Oil	0.500	0.625	0.840	0.967	1.070	1.213	1.421	1.530	1.640	1.842	3.184	3.395	3.986
Value of Products - $/SD	328,325	328,325	328,325	328,325	328,325	328,325	328,325	328,325	328,325	328,325	328,325	328,325	328,325
Less: Cost of Crude	239,144	237,835	235,562	234,204	233,568	232,998	232,520	232,793	232,849	232,889	232,917	232,982	233,114
Capital Related Costs	38,302	37,664	36,640	36,050	35,126	34,374	33,667	33,182	32,930	32,699	32,485	32,396	32,217
Variable Operating Costs	14,047	13,751	13,256	12,961	12,710	12,528	12,472	12,343	12,261	12,398	12,359	12,349	12,328
Gross Realization	36,832	39,075	42,867	45,110	46,921	48,425	49,666	50,007	50,285	50,339	50,564	50,598	50,666
Crude Run, BPSD	98,148	97,616	96,692	96,140	95,881	95,649	95,455	95,566	95,589	95,605	95,617	95,643	95,697
Product Slate, BPSD	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500
Realization, $/Bbl Crude	0.370	0.400	0.443	0.469	0.489	0.506	0.520	0.523	0.526	0.527	0.529	0.529	0.529
Realization, $/Bbl Products	0.402	0.432	0.474	0.498	0.518	0.535	0.549	0.553	0.556	0.556	0.559	0.559	0.560
Crude Cost/Bbl Product	2.642	2.628	2.603	2.588	2.581	2.575	2.569	2.572	2.573	2.573	2.574	2.574	2.576
Cap. & Var. Cost/Bbl Product	0.578	0.570	0.551	0.542	0.529	0.518	0.510	0.503	0.499	0.498	0.496	0.494	0.492
△Crude Cost/Bbl Product	0.066	0.052	0.027	0.012	0.005	- 0.001	- 0.007	- 0.004	- 0.003	-0.003	-0.002	-0.002	---
△Cap. & Var. Cost/Bbl Product	0.086	0.078	0.059	0.050	0.037	0.026	0.018	0.011	0.007	0.006	0.004	0.002	---
Total △ Cost/Bbl Product	0.152	0.130	0.086	0.062	0.042	0.025	0.011	0.007	0.004	0.003	0.002	---	---
Total △Cost/Bbl Fuel Oil	0.544	0.465	0.308	0.222	0.150	0.089	0.039	0.025	0.014	0.011	0.007	---	---
REFINERY CONFIGURATION													
Crude Unit, BPSD	98,148	97,616	96,692	96,140	95,881	95,649	95,455	95,566	95,589	95,605	95,617	95,643	95,697
Reformer Pretreater, BPSD	21,945	21,552	20,961	20,640	19,890	19,778	19,378	19,456	19,361	19,249	19,252	19,270	19,306
Reformer, BPSD	20,981	20,578	19,859	19,420	19,383	19,421	19,251	19,529	19,585	20,432	20,431	20,448	20,481
FCCU Feed Pretreater, BPSD	9,915	9,821	10,849	11,866	4,150	1,752	3,020	0	0	0	0	0	0
FCC Unit, BPSD	18,773	19,636	20,896	21,582	23,414	23,962	23,560	23,077	22,821	20,028	20,021	19,992	19,933
Mid-distillate H_2-Treater, BPSD	24,787	23,516	20,824	19,160	21,245	20,760	14,899	14,523	12,647	12,602	12,644	12,664	12,704
Delayed Coker, BPSD	25,508	23,256	19,324	16,948	16,140	15,383	15,091	15,845	16,104	15,593	15,686	15,823	16,098
Alkylation Unit, BPSD	2,746	2,803	2,911	2,979	2,935	2,891	2,998	2,967	2,979	2,984	2,985	2,981	2,975
Resid, HDS Unit, BPSD	1,430	1,430	1,430	1,402	1,375	1,268	1,220	1,138	1,023	930	562	375	0
Hydrocracking Unit, BPSD	2,693	2,795	2,841	2,820	3,724	3,955	4,270	4,488	4,678	6,080	6,072	6,064	6,048
Hydrogen Plant, MSCFD	2,676	2,741	3,262	3,715	1,025	0	0	0	0	0	0	0	0
Sulfur Plant, M LB PSD	265	256	240	231	216	203	189	175	168	168	168	169	170

CASE NO. 7

EFFECT OF FUEL OIL SULFUR RESTRICTIONS ON AN EXISTING REFINERY

TABLE NO. 10

Machine Iteration No.	17-69	17-71	17-72	17-74	17-83	17-89	17-94	17-98	17-109	17-114	17-115	17-117
Wt. % Sulfur in Fuel Oil	0.500	0.602	0.843	1.082	1.186	1.292	1.499	1.631	1.741	2.124	3.271	3.276
Value of Products - $/SD	328,825	328,825	328,825	328,825	328,825	328,825	328,825	328,825	328,825	328,825	328,825	328,825
Less: Cost of Crude	240,330	239,211	236,533	234,059	233,506	233,722	233,435	233,170	232,969	233,112	233,098	233,097
Capital Related Costs	1,128	1,142	1,155	1,010	612	404	23	339	388	163	---	---
Variable Operating Costs	19,622	18,917	17,287	15,683	15,230	14,184	13,372	12,650	12,519	12,494	12,471	12,456
Gross Realization	67,745	69,555	73,850	78,073	79,477	80,515	81,995	82,666	82,949	83,056	83,256	83,272
Crude Run, BPSD	98,630	98,175	97,087	96,081	95,856	95,944	95,827	95,719	95,638	95,696	95,690	95,690
Product Slate, BPSD	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500
Realization, $/Bbl Crude	0.687	0.708	0.761	0.813	0.829	0.839	0.856	0.864	0.867	0.868	0.870	0.870
Realization, $/Bbl Products	0.749	0.769	0.816	0.863	0.878	0.890	0.906	0.913	0.917	0.918	0.920	0.920
Crude Cost/Bbl Product	2.656	2.643	2.613	2.586	2.580	2.582	2.579	2.576	2.574	2.576	2.575	2.575
Cap. & Var Cost/Bbl Product	0.229	0.222	0.204	0.184	0.175	0.161	0.148	0.144	0.143	0.140	0.138	0.138
Δ Crude Cost/Bbl Product	0.081	0.068	0.038	0.011	0.005	0.007	0.004	0.001	- 0.001	0.001	---	---
Δ Cap. & Var. Cost/Bbl Product	0.091	0.084	0.066	0.046	0.037	0.023	0.019	0.006	0.005	0.002	---	---
Total ΔCost/Bbl Product	0.172	0.152	0.104	0.057	0.042	0.030	0.014	0.007	0.004	0.003	---	---
Total ΔCost/Bbl Fuel Oil	0.615	0.544	0.372	0.204	0.150	0.107	0.050	0.025	0.018	0.011	---	---

REFINERY CONFIGURATION NEW CAPACITY (UNUSED EXISTING CAPACITY IN PARENTHESES)

	Base												
Crude Unit, BPSD	100,000	(1,370)	(1,825)	(2,913)	(3,919)	(4,144)	(4,056)	(4,173)	(4,281)	(4,362)	(4,304)	(4,310)	(4,310)
Reformer Pretreater, BPSD	23,000	(1,050)	(1,444)	(2,361)	(3,133)	(2,905)	(2,320)	(2,377)	(2,509)	(2,530)	(2,514)	(2,514)	(2,518)
Reformer, BPSD	23,000	(755)	(1,179)	(2,167)	(3,037)	(3,048)	(3,135)	(3,382)	(3,631)	(3,969)	(3,911)	(3,946)	(3,926)
FCCU Feed Pretreater, BPSD	---	---	---	---	---	---	---	---	---	---	---	---	---
FCC Unit, BPSD	20,000	378	1,147	2,937	4,292	3,213	248	---	---	---	---	---	---
Mid-Distillate H_2-Treater, BPSD	13,000	13,547	12,447	9,973	8,323	9,361	9,203	4,020	72	---	---	---	(109)
Delayed Coker, BPSD	16,000	10,268	8,377	3,972	---	---	---	---	---	(294)	---	---	---
Alkylation Unit, BPSD	3,000	(664)	(592)	(424)	(259)	(218)	(140)	(90)	(40)	---	---	---	---
Resid. HDS Unit, BPSD	---	1,180	1,180	1,151	1,123	926	611	34	513	587	247	---	---
Hydrocracking Unit, BPSD	7,000	---	---	---	(694)	(1,770)	(1,436)	(355)	(174)	(686)	(751)	(712)	(730)
Hydrogen Plant MSCFD	---	2,162	2,249	2,447	1,163	---	---	---	---	---	---	---	---
Sulfur Plant, M LB PSD	170	90	82	64	45	40	35	25	10	2	2	3	2

CASE NO. 8

TABLE NO. 11

EFFECT OF INCLUDING 5000 BPSD KUWAIT CRUDE AND 500 BPSD OF KUWAIT NO. 6 FUEL - CHEAP NATURAL GAS AVAILABLE

Machine Iteration No.	19-67	19-71	19-72	19-76	19-82	19-89	19-90	19-97	19-104	19-105	19-106	19-108	19-112
Wt. % Sulfur in Fuel Oil	0.500	0.625	0.803	1.061	1.167	1.303	1.451	1.609	1.716	2.838	3.458	3.986	4.103
Value of Products - $/SD	328,825	328,825	328,825	328,825	328,825	328,825	328,825	328,825	328,825	328,825	328,825	328,825	328,825
Less: Cost of Crude	235,818	234,126	232,057	229,067	230,529	230,131	230,096	230,433	230,327	230,200	230,160	230,167	230,170
Capital Related Costs	38,706	38,101	37,253	36,040	34,616	34,008	33,451	32,897	32,599	32,441	32,364	32,206	32,204
Variable Operating Costs	16,624	16,623	16,379	16,012	14,484	14,332	14,087	13,757	13,989	14,058	14,058	14,130	14,129
Gross Realization	37,677	39,975	43,136	47,706	49,196	50,354	51,191	51,738	51,910	52,126	52,243	52,322	52,322
Crude Run, BPSD	96,796	96,108	95,267	94,052	94,646	94,484	94,470	94,607	94,564	94,512	94,496	94,500	94,455
Product Slate, BPSD	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500	90,500
Realization, $/Bbl Crude	0.389	0.415	0.452	0.507	0.519	0.532	0.541	0.546	0.548	0.551	0.552	0.553	0.570
Realization, $/Bbl Products	0.416	0.441	0.476	0.527	0.543	0.556	0.565	0.571	0.573	0.575	0.577	0.578	0.595
Crude Cost/Bbl Product	2.605	2.587	2.564	2.531	2.547	2.542	2.542	2.546	2.545	2.543	2.543	2.543	2.543
Cap. & Var. Cost/Bbl Product	0.611	0.604	0.592	0.575	0.542	0.534	0.525	0.515	0.514	0.513	0.512	0.512	0.512
ΔCrude Cost/Bbl Product	0.063	0.043	0.021	- 0.012	0.004	- 0.001	- 0.001	0.003	0.002	0.000	0.000	0.000	Base
ΔCap. & Var. Cost/Bbl Product	0.099	0.092	0.080	0.063	0.030	0.022	0.013	0.003	0.002	0.001	0.000	0.000	Base
Total ΔCost/Bbl Product	0.162	0.135	0.101	0.051	0.034	0.021	0.012	0.006	0.004	0.001	0.000	0.000	Base
Total ΔCost/Bbl Fuel Oil	0.578	0.488	0.363	0.182	0.123	0.077	0.044	0.023	0.016	0.007	0.003	0.000	Base
REFINERY CONFIGURATION													
Crude Unit, BPSD	96,796	96,108	95,267	94,052	94,646	94,484	94,470	94,607	94,564	94,512	94,496	94,500	94,500
Reformer Pretreater, BPSD	21,717	21,247	20,729	19,978	19,704	19,362	19,214	19,123	19,059	19,031	19,049	19,056	19,056
Reformer, BPSD	22,285	22,077	21,574	20,846	20,702	20,476	20,386	20,377	20,726	20,824	20,853	20,957	20,958
FCCU Feed Pretreater, BPSD	11,066	10,336	10,985	11,910	0	0	0	0	0	0	0	404	404
FCC Unit, BPSD	13,152	13,410	14,144	15,210	19,702	20,024	20,160	20,098	18,897	18,553	18,487	18,035	18,035
Mid-distillate H_2-Treater, BPSD	24,059	22,996	20,964	17,853	22,747	19,859	16,470	13,515	12,589	12,613	12,735	12,561	12,561
Delayed Coker, BPSD	25,816	23,706	20,498	15,878	15,850	15,141	15,095	15,551	15,979	15,960	15,942	16,137	16,137
Alkylation Unit, BPSD	2,882	2,947	3,041	3,178	2,911	2,969	2,993	3,008	2,960	2,944	2,935	2,929	2,929
Resid. HDS Unit, BPSD	1,430	1,430	1,387	1,378	1,300	1,248	1,214	1,032	671	447	332	0	0
Hydrocracking Unit, BPSD	4,887	5,358	5,562	5,858	5,871	6,046	6,110	6,183	6,787	6,982	7,007	7,152	7,152
Hydrogen Plant, MSCFD	6,833	7,138	7,720	8,558	2,052	1,483	980	0	0	0	0	0	0
Sulfur Plant, M LB PSD	267	256	244	224	204	194	183	172	170	171	171	173	173

FIGURES

GENERALIZED REFINERY FLOWSHEET

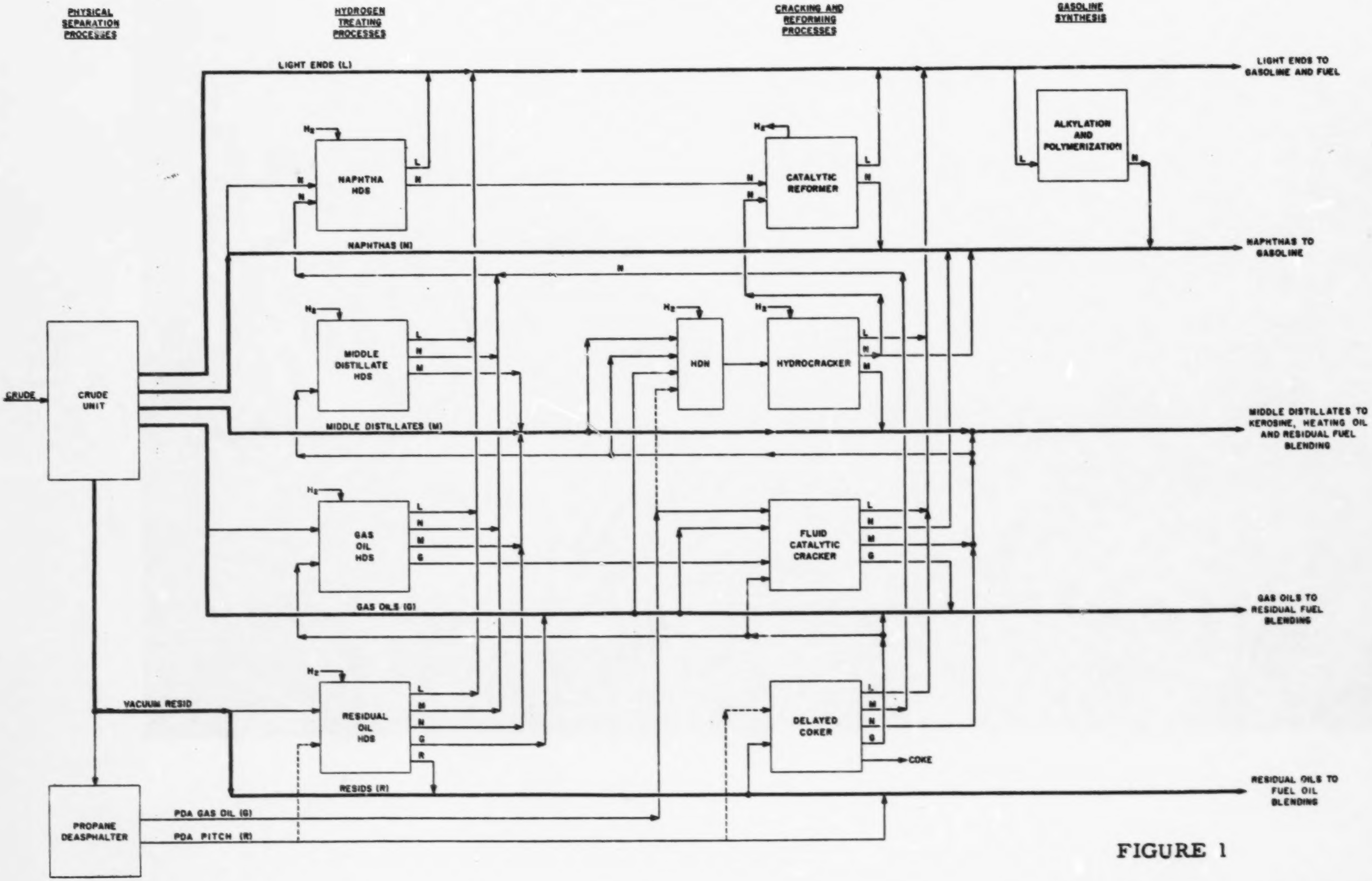

PHYSICAL SEPARATION PROCESSES
HYDROGEN TREATING PROCESSES
CRACKING AND REFORMING PROCESSES
GASOLINE SYNTHESIS
CRUDE
CRUDE UNIT
LIGHT ENDS (L)
NAPHTHAS (N)
MIDDLE DISTILLATES (M)
GAS OILS (G)
VACUUM RESID
RESIDS (R)
PROPANE DEASPHALTER
PDA GAS OIL (G)
PDA PITCH (R)
H_2
NAPHTHA HDS
MIDDLE DISTILLATE HDS
GAS OIL HDS
RESIDUAL OIL HDS
HDN
CATALYTIC REFORMER
HYDROCRACKER
FLUID CATALYTIC CRACKER
DELAYED COKER
COKE
ALKYLATION AND POLYMERIZATION
L
N
M
G
R
LIGHT ENDS TO GASOLINE AND FUEL
NAPHTHAS TO GASOLINE
MIDDLE DISTILLATES TO KEROSINE, HEATING OIL AND RESIDUAL FUEL BLENDING
GAS OILS TO RESIDUAL FUEL BLENDING
RESIDUAL OILS TO FUEL OIL BLENDING

FIGURE 1

FIGURE 2

REFINERY REALIZATION AND
INCREMENTAL COST OF FUEL OIL
RESULTING FROM SULFUR RESTRICTION

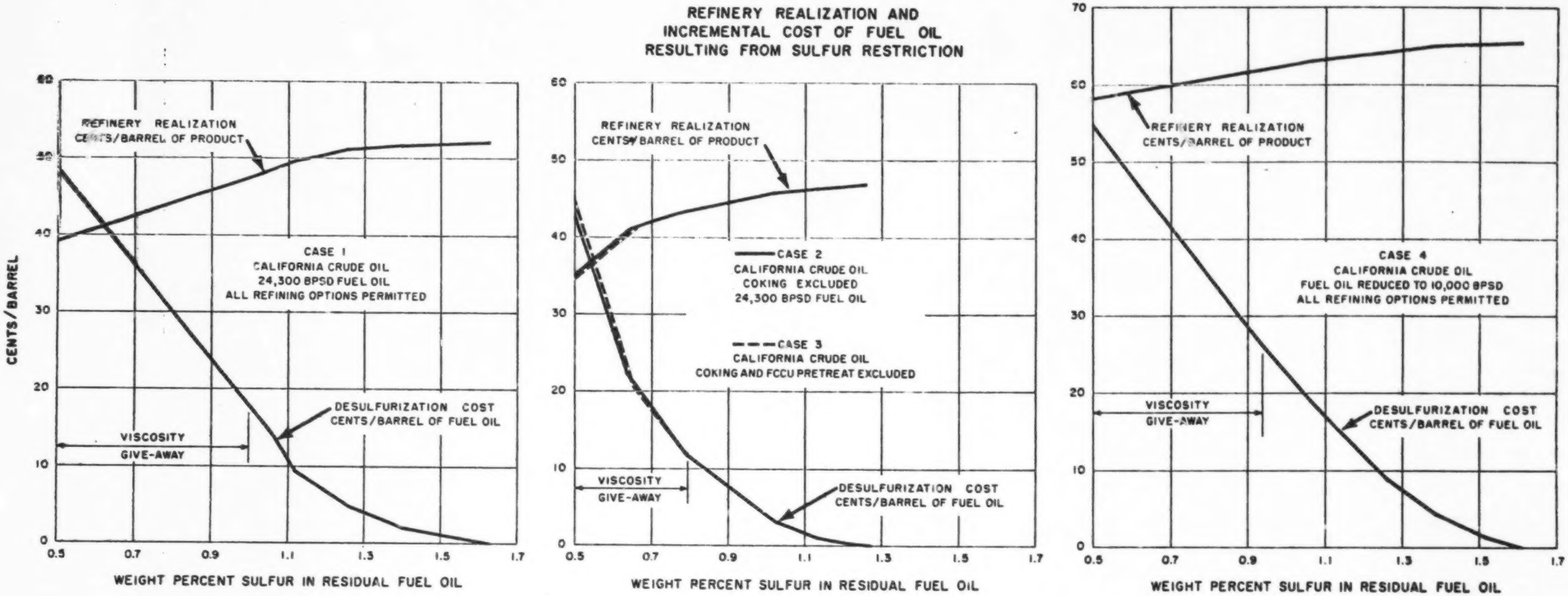


FIGURE 3

REFINERY REALIZATION AND
INCREMENTAL COST OF FUEL OIL
RESULTING FROM SULFUR RESTRICTION

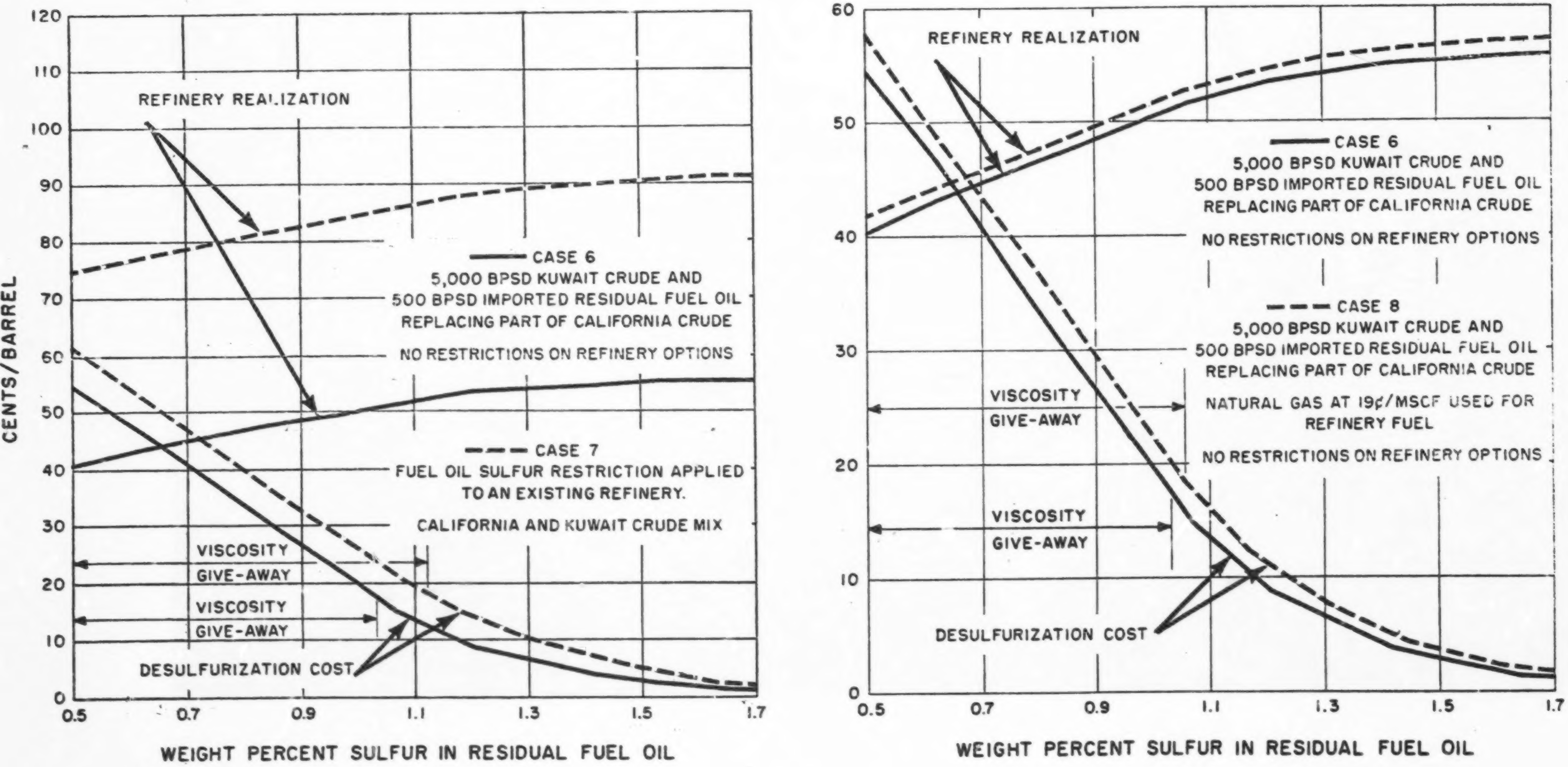


FIGURE 4

CASE 6

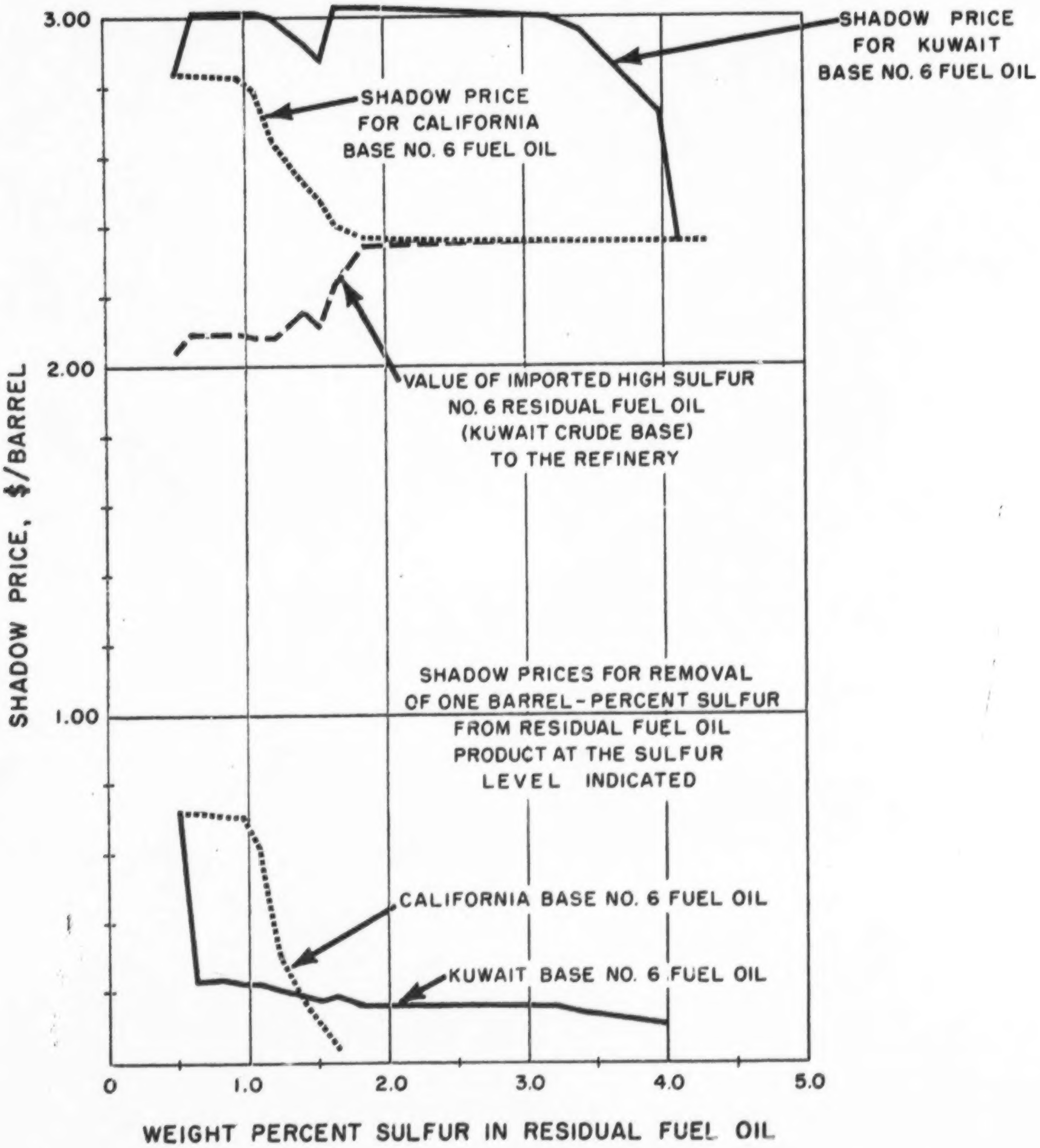


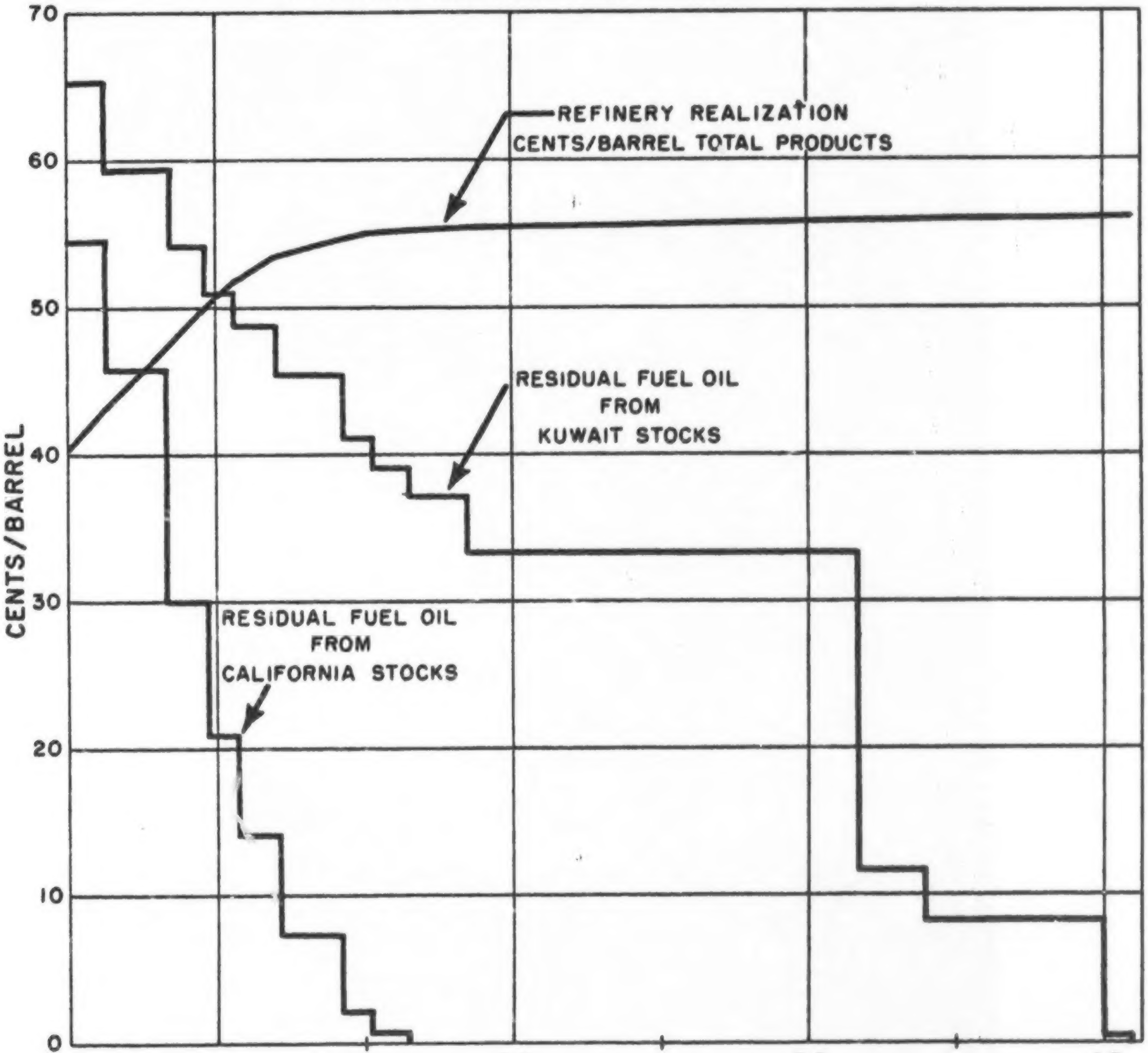
FIGURE 5

CASE 6

COMPARISON OF DESULFURIZATION COST FOR CALIFORNIA AND KUWAIT RESIDUAL FUEL OILS

5000 BPSD KUWAIT CRUDE AND
500 BPSD IMPORTED RESIDUAL FUEL OIL
REPLACING PART OF CALIFORNIA CRUDE

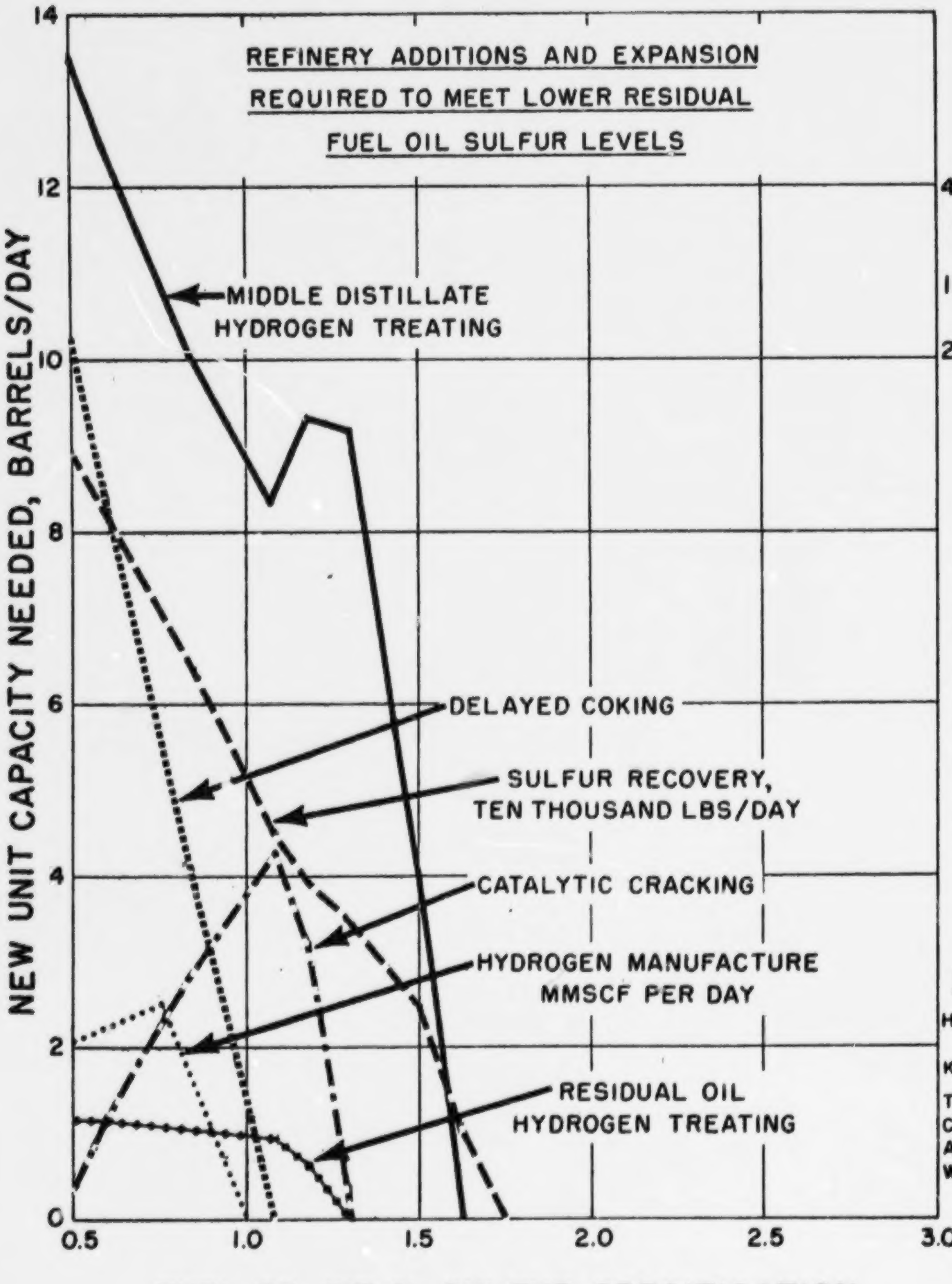
NO RESTRICTIONS ON REFINERY OPTIONS



WEIGHT PERCENT SULFUR IN RESIDUAL FUEL OIL

FIGURE 6

STUDY OF EXISTING 100,000 BARREL REFINERY



CASE 7

CONSTANT PRODUCT DEMANDS

40,300 BPSD GASOLINE
8,000 BPSD LIGHT DISTILLATE
16,900 BPSD HEAVY DISTILLATE
25,300 BPSD RESIDUAL FUEL

BASE REFINERY

UNIT	CAPACITY
CRUDE	100,000 BPSD
FLUID CATALYTIC CRACKER	20,000 "
HYDROCRACKER	7,000 "
MID DIST. HYDROTREATER	13,000 "
CATALYTIC REFORMER	23,000 "
NAPHTHA HYDROTREATER	23,000 "
DELAYED COKER	16,000 "
ALKYLATION	3,000 "
SULFUR RECOVERY	170,000 LBS/D

BASE CASE CRUDE MIX WITH NO SULFUR RESTRICTION

HUNTINGTON BEACH CALIFORNIA	90,680 BPSD
KUWAIT	5,000 BPSD

THE KUWAIT CRUDE RATE WAS HELD CONSTANT AT ALL SULFUR LEVELS AND THE CALIFORNIA CRUDE RATE WAS ALLOWED TO VARY.

APPENDIX

TECHNICAL ASPECTS OF THE LP MATRIX

This appendix presents the data used in the LP matrix. It is believed that no useful purpose would be accomplished by giving a matrix tableau or a listing of the input data cards. The exact arrangement of the matrix and the form of the input data depends on the particular LP code that is being used, and undoubtedly would be modified for any extension of this study. The data in this appendix is sufficiently complete and detailed that the problem can be reconstructed with little effort for any LP code.

Even with modern high speed computers, there are definite economic limitations on the size of the matrices to be solved. The time for solution varies roughly as the cube of the number of equations, so that there is considerable incentive for keeping the matrix small. In this particular problem, a large number of parallel solutions were desired, and the effort involved in creating a compact matrix was justified. The following is a brief description of some of the techniques used:

1. Stream combinations Many of the by-product streams from various refinery units are produced in small amounts. Very little is lost in simply combining these by-product streams with others of similar nature. As an example, the by-product naphthas from the heavy oil hydrodesulfurizer units were all considered to be equivalent to virgin naphtha from crude. This same combining technique was employed for most of the by-product distillates and gas oils from the hydrodesulfurizers which were considered equivalent to the virgin material from crude. One consequence of this is that an exact sulfur balance was not possible. As the sulfur by-product was assigned no value, this discrepancy is of little importance.

2. Non-restricting constraints In most LP problems many of the balance equations do not represent real constraints from a mathematical viewpoint. For example, no restriction was placed on the availability of crude. Practically, its cost can be merged into the cost for the crude unit operations. The Bonner and Moore LP matrix generator has features which facilitate this kind of operation. A large fraction of the equations were made non-restricting in this way.

3. Compositing When two or more units or operations are related by a single feed-product connecting stream, this stream can be eliminated as a constraint by creating a composite operation for each permutation that exists. Here again, the Bonner and Moore code facilitated this operation. The code also includes facilities for reconstructing the results in terms of the original matrix

By applying these techniques, the reduced machine matrix was held to 57 equations and 95 non-basic variables. Actual machine solution time was about 2 minutes for Cases 1 to 5 with approximately an equal amount of time for the report writers. Cases 6 to 8 took slightly longer.

Most of the data presented in this report was generated by the technique of parametric programming. Starting with an optimum solution, the computer program was instructed to explore the effect of a change in sulfur specification until some basic change occurred in the structure of the computer solution. Because 20 or more steps were involved in most cases, printed output was taken only when the step involved a total change of 0. 1 wt. percent sulfur or more. This accounts for the seemingly random selection of sulfur specifications in the tabulations.

The results from the parametric programming steps are particularly instructive in showing the relationships between factors of interest. Part of the printed results from the machine is a tabulation showing how each basic variable reacts to the change in parametrized specification (the sulfur spec.) The rate of hydrogen consumption change with each "barrel-percent" sulfur specification change varied widely with each change in refinery structure. This simply confirmed the observation that there is no basic hydrogen consumption rate that can be related to a given degree of desulfurization of the fuel oil. Hydrogen is simply one of many economic factors that are optimized for total minimum cost.

One particularly useful feature of the LP model was the exact liquid volume balance. Debugging and error checking was greatly facilitated by the fact that the crude run minus the total of the products had to equal a net volume loss calculated by the computer. The volume loss or gain associated with each operation is given in the data listings that follow. These were determined according to the following rules:

1. The volume equivalent fuel oil (6.2 million BTU per barrel) for by-product fuel gas was counted as a product from each operation. The net volume loss or gain is the difference between feed and products for the operation including the fuel gas. Coke was not counted.

2. A separate constraint was set up for refinery fuel. When fuel gas is transferred to this constraint, a volume loss is taken. Similarly, burning product fuel oil, butanes, etc., results in a volume loss.

3. Transferring excess hydrogen to the fuel gas constraint resulted in a volume gain. In turn when this was transferred to refinery fuel, a loss was taken.

In the following tables are listed the various parameters used in the development of the matrix including the following:

Capital Related Charges
Variable Operating Cost Factors
Blending Properties of Product Streams
Process Yields & Utility Demands for each Process
Utility Cost in terms of Fuel
Miscellaneous Refinery Transfers

CAPITAL RELATED CHARGES USED IN LP MODEL

Description of Refinery Units	Base Capacity (Note 1)	Total Investment (Note 2)	Gross Investment Charge ($/Bbl) (Note 3)	Maintenance Cost ($/Bbl) (Note 4)	Insur., Taxes & Misc. Cost ($/Bbl) (Note 4)	Total Capital Related Charge ($/Bbl)
Comb. Crude Unit	80,000 BPSD	$4,700,000	0.0576	0.0069	0.0035	0.068
Propane Deasphalter	20,000 BPSD	$4,000,000	0.1960	0.0264	0.0118	0.234
Reforming Pretreater	20,000 BPSD	$2,200,000	0.1078	0.0146	0.0065	0.129
Catalytic Reformer	20,000 BPSD	$3,300,000	0.1618	0.0219	0.0097	0.193
Delayed Coker	20,000 BPSD	$5,000,000	0.2451	0.0332	0.0147	0.293
Resid. Hydrodesulfurizer	10,000 BPSD	$5,300,000	0.5196	0.1091	0.0312	0.661
Hydrocracker	8,500 BPSD	$4,000,000	0.4614	0.0692	0.0277	0.558
Mid. Dist. H_2 Treater	20,000 BPSD	$2,400,000	0.1176	0.0159	0.0071	0.141
FCCU Feed Pretreater	20,000 BPSD	$3,251,000	0.1593	0.0216	0.0096	0.191
Fluid Catalytic Cracker	20,000 BPSD (5)	$5,800,000	0.2843	0.0383	0.0170	0.340 (5)
Alkylation Unit	4,000 BPSD	$2,600,000	0.6373	0.0860	0.0382	0.762
Sulfur Plant	2,000 M Lbs/SD	$2,700,000	0.0013/M Lbs.	0.0002/M Lbs.	0.0001/M Lbs.	0.002/M Lbs.
Hydrogen Plant	32,000 MSCF/SD	$4,300,000	0.1309/MSCF	0.0227/MSCF	0.0079/MSCF	0.162/MSCF

(1) Nominal stream day unit charge rates assumed for estimating capital investment.
(2) Battery limits construction costs.
(3) Based on 1020 operating days (3 years) pay-out period before taxes, interest and amortization.
(4) Estimated annual cost for these items divided by the total feed for 340 operating days.
(5) Variable capacity factors were used in the LP model to account for difference in coke yield of different feedstocks.

VARIABLE OPERATING COST FACTORS FOR LP MODEL

Operation	Catalyst			Water (3)			Chemicals & Misc.	
	$/lb	lbs/B	$/B	Type	Gal/B	$/B	$/B	Total
FCCU (600-950° gas oil, treated or untreated)	0.1575	0.205	0.0323	BFW	10.3	0.0072		$0.0395
FCCU (700-950° gas oil, treated or untreated)	0.1575	0.205	0.0323	BFW	8.91	0.0063		$0.0386
FCCU (Heavy coker gas oil blended, treated or untreated)	0.1575	0.205	0.0323	BFW	11.2	0.0078		$0.0401
FCCU (PDA gas oil blend, treated or untreated)	0.1575	0.205	0.0323	BFW	12.1	0.0085		$0.0408
FCCU (Increase conversion from 62% to 64.5%)				BFW	0.289	0.000202		$0.000202
Hydrocracker (Virgin Stocks, Max. dist. production)			0.068				0.050	$0.118
Hydrocracker (Virgin Stocks, Max. gasoline Production)			0.137				0.050	$0.187
Hydrocracker (FCCU LCO/coker LGO, max. gasoline production)			0.146				0.050	$0.196
Crude Unit (Atm. & Vac.)							0.005	$0.005
Propane Deasphalter							0.010	$0.010
Catalytic Reformer, (S. R. Gasoline 98 RON reformate)	4.0	0.0222	0.0888					$0.0888
Catalytic Reformer, (95 & 92 RON reformate)	4.0	0.0111	0.0444					$0.0444
Catalytic Reformer, (Hydrocrackate, 98 RON Reformate)	4.0	0.0111	0.0444					$0.0444
Coking, Delayed							0.010	$0.010
Residual Oil Hydrodesulfurizer	0.95	0.00832	0.0792				0.040	$0.0832
Hydrogen Plant (Values per MSCF Hydrogen)			0.0025	BFW	1.785	0.0012		$0.0037
Sulfur Plant (Values per lb. of elemental Sulfur)							0.00175	$0.00175
Alkylation, HF							0.100	$0.100
Reformer Pretreater			0.010					$0.010
Mid-Distillate Hydrotreater	0.75	0.005	0.0038					$0.0038
FCCU Feed Pretreater			0.010					$0.010
Cooling Water Facilities (Values per 1000 gal. Cooling Water)			0.0277(1)	CW	0.0438(4)	0.0044	0.0038	$0.0359
Boiler & Steam Plant (Values per 1000 lbs. steam)				BFW	0.12	0.084		$0.084
Electric Power (Values per KWH)								$0.003(2)
Amine Plant (Values per lb. Sulfur extracted)		–	0.002036(1)					$0.002036

(1) Estimated total including an allowance for maintenance and incremental investment
(2) Electrical cost estimated as $0.003/KWH + 9000 BTU/KWH fuel consumption
(3) Boiler feedwater valued at $0.70/Mgal, Cooling tower make-up valued at $0.10/Mgal.
(4) Cooling tower water make-up ratio, Mgal/Mgal.

BLENDING PROPERTIES OF PRODUCT STREAMS

Blending Component		Research Octane No. With 1cc Pb	2cc Pb	3cc Pb	RVP (PSIA)
Iso-Butane (Note 1)		98.2	100.0	101.4	80.0
N - Butane		98.2	100.0	101.4	59.1
Butylenes		97.2	99.0	100.0	60.0
Reformate	92 RON	95.7	97.7	99.0	5.0
Reformate	95 RON	98.6	99.9	100.8	5.0
Reformate	98 RON	100.6	101.8	102.6	6.0
Saturated C_5-175°		88.5	91.9	93.8	5.4
Coker Light Gasoline		86.1	89.4	91.3	10.5
Light Hydrocrackate		89.2	93.3	95.6	12.0
400 EP Cat. Gasoline		95.6	97.3	98.4	7.1
330 EP S. R. Naphtha		78.6	82.5	84.8	3.0
C_4 Alkylate		100.6	102.5	103.7	5.0
C_3 Alkylate		97.6	99.5	100.7	5.0
Polymer Gasoline		96.6	98.1	99.0	1.0
Kuwait Light Gasoline		84.6	88.6	91.0	12.0
Kuwait Reformate		94.8	96.6	97.7	4.00

(1) Octanes assumed to be same as N-butane to avoid preferential use of iso-butane for gasoline blending.

Blending Component	Viscosity Blending Number	Sulfur (Wt. %)	° API
330/400 - 540 Kerosine	-100	0.52	36.8
540-650 Hvy. Vir. Dist.	+ 30	1.22	30.3
400-650 Raw Coker Dist.	0	1.25	35.0
600-950 FCCU Feed	+390	1.47	21.7
700-950 FCCU Feed	+475	1.51	19.6
650-800 Hydrocracker Feed	+335	1.42	22.9
800-900 Gas Oil	+520	1.54	18.6
Vacuum Pitch, 950+	+1110	2.03	6.2
Propane Deasphalter Pitch	+1750	2.57	-5.1
Resid. HDS Pitch	+1750	1.70	-7.4
Light FCCU Cycle Oil	-5	1.15	23.6
Hvy. FCCU Cycle Oil	+310	1.30	16.5
540-680 H-Oil	+210	0.50	25.8
Hydrocracked Distillate	-100	0	38.0
FCCU Decant Oil	+875	1.40	3.8
Desulf. FCCU Cycle Oil	-5	0.12	24.6
Desulf. 400-650 Coker Dist.	0	0.13	36.0
Desulf. 540-650 Hvy. Vir. Dist.	+ 30	0.12	31.3
Desulf. Kerosine	-100	0.05	37.8
Light Cycle Oil, Trtd. Feed	-5	0.01	26.5
Hvy. Cycle Oil, Trtd. Feed	+310	0.02	24.0
Decant Oil, Trtd. Feed	+875	0.05	10.2
Coker HGO, 650-950	+350	1.55	20.0
380-540 Kuwait Distill.	-150	0.61	41.6
540-650 Kuwait Distill.	+ 40	1.75	32.9
Kuwait FCC & Coker Cycle	- 53	2.40	25.0
Kuwait Heavy Cycle Oil	+334	3.20	15.4
Kuwait Low S. Fr. Distill.	+ 50	0	33.1
Kuwait Desulf. Resid.	+700	0.75	10.0
Kuwait Vacuum Resid.	+870	5.40	3.5
Treated Kuwait Lt. Distill.	-150	0.061	41.6
Treated Kuwait Hvy. Distill.	+ 40	0.175	32.9
Treated KW Coker & Cycle	- 53	0.24	25.0
Kuwait 650-1100	+380	3.00	21.2
Kuwait Coker Heavy G. O.	+400	4.00	18.0
Imported Fuel Oil	569	4.50	12.0

CRUDE DISTILLATION

CRUDE OIL DISTILLATION - BASE CASE FOR 330 EP GASOLINE AND HYDROCRACKER FEED GAS OIL

FEED CONSTR.	PRODUCT STREAMS	
1.0		HUNTINGTON BEACH CRUDE
	0.001	ISO-BUTANE
	0.001	NORMAL BUTANE
	0.021	PENTANES-175
	0.125	175-330 NAPHTHA, SR
	0.195	LT DISTILLATE 540 EP
	0.102	HVY DISTILLATE 650 EP
	0.125	HYDROCRACKER FEED 650-800
	0.125	HEAVY GAS OIL 800-950
	0.305	VACUUM PITCH 950+
0.0		VOLUME GAIN
1.0		CAPACITY CONSTRAINT FACTOR
0.600		WATER, MGAL/B
0.025		STEAM, MLBS/B
0.013		FUEL, FOE/B
0.400		ELECTRICAL, KWH/B
	0.064	GASOLINE/ DIST SWING

CRUDE DISTILLATION - 330 EP GASOLINE AND FULL RANGE FCCU FEED

FEED CONSTR.	PRODUCT STREAMS	
1.0		HUNTINGTON BEACH CRUDE
	0.001	ISOBUTANE
	0.001	NORMAL BUTANE
	0.021	PENTANES-175
	0.125	175-330 NAPHTHA, SR
	0.195	LT DISTILLATE 540 EPP
	0.057	HVY DISTILLATE 600 EP
	0.295	FULL RANGE CAT FEED 600-950
	0.305	VACUUM PITCH 950+
0.0		VOLUME GAIN
1.0		CAPACITY CONSTRAINT FACTOR
0.600		WATER MGAL/B
0.025		STEAM MLBS/B
0.013		FUEL FOE/B
0.400		ELECTRICAL KWH/B
	0.064	GASOLINE/DIST SWING

CRUDE DISTILLATION - 330 EP GASO 'AND 700-950 FCCU FEED

FEED CONSTR.	PRODUCT STREAMS	
1.0		HUNTINGTON BEACH CRUDE
	0.001	ISOBUTANE
	0.001	NORMAL BUTANE
	0.021	PENTANES-175
	0.125	REFORMER FEED 330 EP
	0.195	LT DISTILLATE 540 EP
	0.142	HVY DISTILLATE 700 EP
	0.210	HIGH BP CAT FEED 700-950
	0.305	VACUUM PITCH 950+
0.0		VOLUME GAIN
1.0		CAPACITY CONSTRAINT FACTOR
0.600		WATER MGAL/B
0.025		STEAM MLBS/B
0.013		FUEL FOE/B
0.400		ELECTRICAL KWH/B
	0.064	GASOLINE/DIST SWING

SWING DISTILLATE TO GASOLINE

FEED CONSTR.	PRODUCT STREAMS	
0.064		LIGHT DISTILLATE 540 EP
0.125		175-330 NAPHTHA, SR
	0.189	175-400 NAPHTHA, SR
0.0		VOLUME GAIN
0.064		LIMIT DIST/GASO SWING

NOTE: COLUMN LABELED "FEED CONSTRAINTS"

REPRESENTS POSITIVE MATRIX COEFFICIENTS

COLUMN LABELED "PRODUCT STREAMS"

REPRESENTS NEGATIVE MATRIX COEFFICIENTS.

VACUUM RESIDUE OPERATIONS

RESIDUAL OIL HYDRO-DESULFURIZATION

FEED CONSTR.	PRODUCT STREAMS	
1.0		VACUUM PITCH 950+
1.000		HYDROGEN,MSCF/B
	0.131	400-540 CUT
	0.130	540-680 CUT
	0.369	680-975 FCCU CUT
	0.200	H-OIL RESID
	0.058	C1-C3,FOE
	4.000	SULFUR AS H2S,LBS/BBL
	0.01	N-BUTANE
	0.07	C5-175 GASOLINE
	0.13	175-400 GASOLINE
0.098		VOLUME GAIN
1.0		CAPACITY CONSTRAINT FACTOR
0.0312		STEAM, MLBS/B
0.965		COOLING WATER, MGAL/B
8.4		ELECTRICITY, KWH/B
0.0218		FUEL FOE B/B

DELAYED COKING

FEED CONSTR.	PRODUCT STREAMS	
1.0		VACUUM PITCH 950+
	3.13	SULFUR IN GAS,LBS/BBL,
	0.1081	FUEL GAS (FOE)
	0.0124	PROPYLENE
	0.0055	ISO-BUTANE
	0.0163	BUTYLENE
	0.0162	NORMAL BUTANE
	0.0771	COKER LT GASOLINE
	0.1349	COKER HVY GASOLINE 200-400
	0.3636	COKER LGO 400-650
	0.1414	COKER HGO 650-950
	0.1107	DELAYED COKE PRODUCT,MLBS/B
	0.1245	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.038		STEAM, MLBS/B
1.440		ELECTRICITY, KWH/B
0.0329		PROCESS FUEL, FOE/B
1.300		COOLING WATER, MGAL/B

PROPANE DE-ASPHALTING

FEED CONSTR.	PRODUCT STREAMS	
1.0		VACUUM PITCH 950+
	0.528	DECARBONIZED OIL
	0.472	PROPANE DEASPHALTED PITCH
	0.0	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.050		STEAM MLBS/B
0.0532		PROCESS FUEL, FOE/B
1.500		COOLING WATER MGAL/B

REFORMER FEED PRETREATING

REFORMER FEED PRETREATING, VIRGIN

FEED CONSTR.	PRODUCT STREAMS	
1.0		330 EP VIRGIN NAPHTHA RAW
0.050		HYDROGEN, MSCF/B
	0.995	330 EP VIRGIN NAPHTHA TRTD
	0.24	H2S, SULFUR LBS/BBL
	0.005	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.013		PROCESS FUEL, FOE/B
0.75		ELECTRICITY, KWH/B

REFORMER FEED PRETREATING
400 EP NAPHTHA

FEED CONSTR.	PRODUCT STREAMS	
1.0		400 EP VIRGIN NAPHTHA RAW
0.050		HYDROGEN, MSCF/B
	0.995	400 EP VIRGIN NAPHTHA TRTD
	0.39	H2S, SULFUR LBS/B
	0.005	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.013		FUEL OIL FOE B/B
0.75		ELECTRICITY KWH/B

REFORMER FEED PRETREATING, COKER
HEAVY NAPHTHA

FEED CONSTR.	PRODUCT STREAMS	
1.0		COKER HEAVY GASOLINE
0.120		HYDROGEN, MSCF/B
	0.995	PRETREATED CRACKED NAPHTHA
	0.68	H2S, SULFUR LBS/BBL
	0.005	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.013		FUEL OIL BBL/BBL
0.75		ELECTRICITY, KWH/B

CATALYTIC REFORMING

REFORM HEAVY HYDROCRACKER GASOLINE
TO 98 RON

FEED CONSTR.	PRODUCT STREAMS	
1.0		HEAVY HYDROCRACKATE
	0.7890	HYDROGEN MSCF/B
	0.0324	FOE
	0.0132	I-C4
	0.0212	N-C4
	0.9110	REFORMATE, 98
	0.0222	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.0647		FUEL FOE
0.0010		STEAM MLBS/B
3.8		ELECTR KWH/B
0.745		COOLING WATER MGAL/B

REFORMER COKER NAPHTHA 92 ON

FEED CONSTR.	PRODUCT STREAMS	
1.0		COKER HVY GASOLINE
	0.6500	HYDROGEN MSCF/BBL
	0.0723	FOE
	0.0277	ISO-BUTANE
	0.0443	N-BUTANE
	0.8260	REFORMATE, 92
	0.0297	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.0647		FUEL, FOE
0.0010		STEAM MLBS/B
3.8		ELECTRICITY KWH/B
0.745		COOLING WATER MGAL/B

CATALYTIC REFORMING

REFORMING 330 EP NAPHTHA TO 92 RON
CLEAR

FEED CONSTR.	PRODUCT STREAMS	
1.0		175-330 NAPHTHA, SR
	0.925	HYDROGEN,MSCF/B
	0.0202	FUEL GAS, FOE
	0.0086	ISO BUTANE
	0.0136	NORMAL BUTANE
	0.9260	REFORMATE, 92RON
	0.0316	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.0647		FUEL, FOE
0.0010		STEAM, MLBS/B
3.8		ELECTRICITY, KWH/B
0.745		COOLING WATER, MGAL/B

REFORMING 330 EP VIRGIN NAPHTHA
TO 98 RON CLEAR

FEED CONSTR.	PRODUCT STREAMS	
1.0		175-330 NAPHTHA SR
	1.0000	HYDROGEN MSCF/B
	0.0394	FUEL GAS,FOE
	0.0159	ISO-BUTANE
	0.0255	NORMAL BUTANE
	0.8740	REFORMATE, 98 OCTANE
	0.0452	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.0647		FUEL,FOE
0.0010		STEAM, MLBS/B
3.8		ELECTR. KWH/B
0.745		COOLING WATER, MGAL/B

REFORMING AT 95 RON CLEAR, 330
REFORMING 330 EP VIRGIN NAPHTHA
TO 95 RON CLEAR

FEED CONSTR.	PRODUCT STREAMS	
1.0		175-330 NAPHTHA, SR
	0.9750	HYDROGEN, MSCF/B
	0.0364	FUEL GAS, FOE
	0.0146	ISO-BUTANE
	0.0232	NORMAL BUTANE
	0.8820	REFORMATE, 95 RON
	0.0438	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.0647		FUEL FOE
0.0010		STEAM MLBS/B
3.8		ELECTR. KWH/B
0.745		COOLING WATER MGAL/B

CATALYTIC REFORMING

REFORMING 400 EP VIRGIN NAPHTHA
TO 92 RON CLEAR

FEED CONSTR.	PRODUCT STREAMS	
1.0		175-400 NAPHTHA, SR
	0.915	HYDROGEN,MSCF/B
	0.0285	FUEL GAS,FOE
	0.0120	ISO BUTANE
	0.0188	NORMAL BUTANE
	0.9120	REFORMATE, 92RON
	0.0287	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.0647		FUEL, FOE
0.0010		STEAM, MLBS/B
3.8		ELECTRICITY, KWH/B
0.745		COOLING WATER, MLBS/B

REFORMING 400 EP VIRGIN NAPHTHA
TO 95 RON CLEAR

FEED CONSTR.	PRODUCT STREAMS	
1.0		175-400 NAPHTHA, SR
	0.9680	HYDROGEN MSCF/B
	0.0296	FOE
	0.0120	I-BUTANE
	0.0193	N-BUTANE
	0.9065	REFORMATE, 95 RON
	0.0326	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.0647		FUEL,FOE
0.0010		STEAM MLBS/B
3.8		ELECTR. KWH/B
0.745		COOLING WATER MGAL/B

REFORMING 400 EP VIRGIN NAPHTHA TO
98 RON CLEAR

FEED CONSTR.	PRODUCT STREAMS	
1.0		175-400 NAPHTHA
	0.9900	HYDROGEN, MSCF/B
	0.0319	FOE
	0.0130	I-C4
	0.0208	N-BUTANE
	0.8990	REFORMATE 98 ON
	0.0353	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.0647		FUEL, FOE
0.0010		STEAM MLBS/B
3.8		ELECTRICITY KWH/B
0.745		COOLING WATER MGAL/B

MIDDLE DISTILLATE HYDROGEN TREATING

MIDDLE DIST HYDROGEN TREATER
LIGHT VIRGIN DISTILLATES

FEED CONSTR.	PRODUCT STREAMS	
1.0		MIDDLE DISTILLATE 350-560V
0.100		HYDROGEN,MSCF/B,
	0.013	C5+ GASOLINE
	0.986	MID DIST HTR LVGO PRODUCT
	0.002	N-BUTANE
	0.0110	FUEL GAS FOE
0.012		VOLUME GAIN
1.0		CAPACITY CONSTRAINT FACTOR
0.017		STEAM,MLBS/B
0.33		COOLING WATER MGAL/BBL
1.9		ELECTRICITY KWH/B
0.0183		PROCESS FUEL, FOE/B

MIDDLE DISTILLATE HYDROGEN TREATER
HEAVY VIRGIN DISTILLATES

FEED CONSTR.	PRODUCT STREAMS	
1.0		MID DIST 560-650 V
0.200		HYDROGEN,MSCF/B
	0.028	C5+ GASOLINE
	0.970	MID DIST HTR HVGO PRODUCT
	0.005	N-BUTANE
	0.0134	FUEL GAS FOE
0.0164		VOLUME GAIN
1.0		CAPACITY CONSTRAINT FACTOR
0.017		STEAM, MLBS/B 170PSIG
0.33		COOLING WATER MGAL/BBL
1.9		ELECTRICITY KWH/B
0.0183		PROCESS FUEL, FOE/B

MIDDLE DISTILLATE HYDROGEN TREATER
COKER DISTILLATE AND FCCU LIGHT CYCLE OIL

FEED CONSTR.	PRODUCT STREAMS	
1.0		MIDDLE DISTILLATE 400-650LC
0.300		HYDROGEN,MSCF/B
	0.041	C5+ GASOLINE
	0.967	MID DIST HTR LCO PRODUCT
	0.004	N-BUTANE
	0.0134	FUEL GAS FOE
0.0254		VOLUME GAIN
1.0		CAPACITY CONSTRAINT FACTOR
0.017		STEAM, MLBS/B 170PSIG
0.33		COOLING WATER MGAL/BBL
1.9		ELECTRICITY KWH/B
0.0183		PROCESS FUEL, FOE/B

NOTE: THESE COEFFICIENTS WERE USED BY THE MATRIX GENERATOR TO CREATE A SEPARA "DESULFURIZER" FOR EACH STREAM TO BE TREATED. H_2S YIELDS FROM THESE OPER TIONS WERE ENTERED SEPARATELY.

HYDROCRACKING

HYDROCRACKER - CONVERT VIRGIN LIGHT DISTILLATE TO GASOLINE

FEED CONSTR.	PRODUCT STREAMS		
1.0		540-650 VIRGIN DIST	
1.9		H2 MSCF/B	
	3.74	H2S SULFUR LBS/B	
	0.034	FUEL GAS FOE	
	0.144	ISO BUTANE	
	0.060	N-BUTANE	
	0.364	LT HYDROCRACKATE	C5-180
	0.666	HY HYDROCRACKATE	400EP
0.268		VOLUME GAIN	
1.0		CAPACITY CONSTRAINT FACTOR	
0.0283		STEAM MLBS/B	
1.18		COOLING WATER MGAL/B	
10.0		ELECTRICITY KWH/B	
0.098		FUEL FOE B/B	

HYDROCRACKER - CONVERT 650-800 VIRGIN GAS OIL TO GASOLINE

FEED CONSTR.	PRODUCT STREAMS		
1.0		HYDROCRACKER FEED 650-800	
2.05		H2 MSCF/B	
	4.5	H2S, SULFUR LBS/BBL	
	0.048	FUEL GAS FOE	
	0.152	ISO-BUTANE	
	0.063	N-BUTANE	
	0.382	LT HYDROCRACKATE	C5-180
	0.692	HY HYDROCRACKATE	400EP
0.337		VOLUME GAIN	
1.0		CAPACITY CONSTRAINT FACTOR	
1.18		COOLING WATER MGAL/B	
0.0283		STEAM MLBS/B	
12.6		ELECTRICITY KWH/B	
0.098		FUEL FOE B/B	

HYDROCRACKER 650-800 VIRGIN GAS OIL FEED FOR MAXIMUM LIGHT DISTILLATE PRODUCTION

FEED CONSTR.	PRODUCT STREAMS		
1.0		HYDROCRACKER FEED 650-800	
1.65		H2 MSCF/BBL	
	4.5	H2S, SULFUR LBS/B	
	0.030	FUEL GAS FOE	
	0.016	ISO-BUTANE	
	0.025	N-BUTANE	
	0.061	LT HYDROCRACKATE	C5-180
	0.209	HY HYDROCRACKATE	400EP
	0.815	HYDROCRACKED DISTILLATE	
0.156		VOLUME GAIN	
1.0		CAPACITY CONSTRAINT FACTOR	
0.849		COOLING WATER MGAL/B	
0.0283		STEAM MLBS/B	
8.95		ELECTRICITY KWH/B	
0.0475		FUEL FOE B/B	

HYDROCRACKING

HYDROCRACKER - LIGHT COKER GAS OIL FEED FOR MAXIMUM GASOLINE

FEED CONSTR.	PRODUCT STREAMS		
1.0		COKER LGO 400-650	
3.1		H2 MSCF/B	
	4.2	H2S SULFUR LBS/B	
	0.042	FUEL GAS FOE	
	0.117	ISO-BUTANE	
	0.052	N-BUTANE	
	0.317	LT HYDROCRACKATE	C5-180
	0.780	HY HYDROCRACKATE	400EP
0.308		VOLUME GAIN	
1.0		CAPACITY CONSTRAINT FACTOR	
1.180		COOLING WATER MGAL/B	
0.0283		STEAM MLBS/B	
14.0		ELECTRICITY KWH/B	
0.098		FUEL FOE B/B	

HYDROCRACKER - CONVERT CATALYTIC LIGHT CYCLE OIL TO GASCLINE

FEED CONSTR.	PRODUCT STREAMS		
1.0		LT CYCLE OIL	
3.3		H2, MSCF/B	
	3.9	H2S, SULFUR LBS/BBL	
	0.042	FUEL GAS,FOE	
	0.117	ISO-BUTANE	
	0.052	N-BUTANE	
	0.317	LT HYDROCRACKATE	C5-180
	0.780	HY HYDROCRACKATE	400EP
0.308		VOLUME GAIN	
1.0		CAPACITY CONSTRAINT FACTOR	
0.0283		STEAM MLBS/B	
1.180		COOLING WATER MGAL/B	
14.0		ELECTRICITY KWH/B	
0.098		FUEL FOE B/B	

FLUID CATALYTIC CRACKING UNIT - UNTREATED FEED

FCCU 600-950 GAS OIL 62 CONV.

FEED CONSTR.	PRODUCT STREAMS	
1.0		600- 950 FCCU FEED
0.02872		FCCU REGEN COKE MLBS/BBL
	2.140	H2S, LBS SULFUR/ BBL
	0.0836	FUEL GAS, FOE
	0.0835	PROPYLENE
	0.0480	ISO-BUTANE
	0.0100	N- BUTANE
	0.0840	BUTYLENE
	0.389	400 EP GASOLINE
	0.322	LT CYCLE OIL 400/650
	0.038	HVY.CYCLE OIL
	0.020	DECANT OIL
0.0781		VOLUME GAIN
1.0		CAPACITY CONSTRAINT FACTOR
	0.10820	STEAM PRODUCTION, MLBS/B
0.96200		COOLING WATER, MGAL/B
0.310		ELECTRICITY, KWH/B
0.00292		FUEL FOE, B/B

CATALYTIC CRACKING OF PROPANE DE-ASPHALTED GAS OIL IN BLEND WITH VIRGIN GAS OIL

FEED CONSTR.	PRODUCT STREAMS	
0.3531		PDA GAS OIL
0.6469		600-950 FCCU FEED
0.0330		FCCU REGEN COKE, MLBS/B
	2.46	H2S, LBS SULFUR/B
	0.0885	FUEL GAS, FOE
	0.0806	PROPYLENE
	0.0340	ISO-BUTANE
	0.0090	N-BUTANE
	0.0840	BUTYLENE
	0.3850	400 EP GASOLINE
	0.3450	LT CYCLE OIL
	0.0150	HVY CYCLE OIL
	0.0200	DECANT OIL
0.0611		VOLUME GAIN
1.14903		CAPACITY CONSTRAINT FACTOR
	0.12170	STEAM PRODUCTION, MLBS/B
1.0460		COOLING WATER, MGAL/B
0.392		ELECTRICITY, KWH/B
0.00355		FUEL FOE B/B

FCCU 700-950 GAS OIL 62 CONV.

FEED CONSTR.	PRODUCT STREAMS	
1.0		700-950 FCCU FEED
0.03121		FCCU REGEN COKE MLBS/BBL
	2.47	H2S, LBS SULFUR/BBL
	0.0822	FUEL GAS,FOE
	0.0807	PROPYLENE
	0.0400	ISO-BUTANE
	0.0100	N -BUTANE
	0.0810	BUTYLENE
	0.3960	400 EP GASOLINE
	0.277	LCO
	0.083	HCO
	0.020	DECANT OIL
0.0699		VOLUME GAIN
1.08670		CAPACITY CONSTRAINT FACTOR
	0.09420	STEAM PRODUCTION, MLBS/B
0.96000		COOLING WATER, MGAL/B
0.340		ELECTRICITY, KWH/B
0.00318		FUEL FOE B/B

CATALYTICALLY CRACK HEAVY COKER GAS OIL IN BLEND WITH VIRGIN GAS OIL

FEED CONSTR.	PRODUCT STREAMS	
0.1272		COKER HGO 650 - 950
0.8728		600 - 950 FCCU FEED
0.02968		FCCU REGEN COKE, MLBS/B
	2.44	H2S, LBS SULFUR/B
	0.0826	FUEL GAS, FOE
	0.0836	PROPYLENE
	0.0450	ISO-BUTANE
	0.0100	N-BUTANE
	0.0830	BUTYLENE
	0.3880	400 EP GASOLINE
	0.3220	LT CYCLE OIL
	0.0380	HVY CYCLE OIL
	0.0200	DECANT OIL
0.0722		VOLUME GAIN
1.03343		CAPACITY CONSTRAINT FACTOR
	0.11320	STEAM PRODUCTION, MLBS/B
0.96700		COOLING WATER, MGAL/B
0.322		ELECTRICITY, KWH/B
0.00296		FUEL FOE B/B

HYDROGEN TREATING OF FCCU FEED

HYDROGEN TREAT FCCU FEED, 600-950 VIRGIN GAS OIL

FEED CONSTR.	PRODUCT STREAMS	
1.0		600-950 VIRGIN GAS OIL
0.675		HYDROGEN,MSCF/B
	4.5	H2S AS SULFUR,LBS/B
	0.0286	FUEL GAS FOE
	0.026	N-BUTANE
	0.097	GASOLINE, 48 API
	0.123	DIESEL, 30 API-DESULF
	0.800	FCCU FEED, 600-950 TREATED
0.0746		VOLUME GAIN
1.0		CAPACITY CONSTRAINT FACTOR
0.017		STEAM, MLBS/B
0.50		COOLING WATER, MGAL/B
3.0		ELECTRICITY, KWH/B
0.0183		FUEL, FOE

HYDROGEN TREAT 700-950 VIRGIN GAS OIL FCCU FEED

FEED CONSTR.	PRODUCT STREAMS	
1.0		700-950 VIRGIN GAS OIL
0.675		HYDROGEN, MSCF/B
	4.7	H2S AS SULFUR,LBS/B
	0.0252	FUEL GAS,FOE
	0.0227	N-BUTANE
	0.007	GASOLINE, 48 API
	0.363	DIESEL, 30 API
	0.650	FCCU FEED 700-950 TREATED
0.0679		VOLUME GAIN
1.0		CAPACITY CONSTRAINT FACTOR
0.017		STM, MLBS/B
0.50		COOLING WATER, MGAL/B
3.0		ELECTRICITY, KWH/B
0.0183		FUEL, FOE

HYDROGEN TKEAT PDA GAS OIL FCCU FEED

FEED CONSTR.	PRODUCT STREAMS	
1.0		PDA GAS OIL
0.675		HYDROGEN, MSCF/B
	4.5	H2S AS SULFUR, LBS/B
	0.0286	FUEL GAS, FOE
	0.026	N-BUTANE
	0.097	GASOLINE, 48 API
	0.123	DIESEL, 30 API
	0.800	FCCU FEED PDA GAS OIL TRTD
0.0746		VOLUME GAIN
1.0		CAPACITY CONSTRAINT FACTOR
0.017		STM MLBS/B
0.50		COOLING WATER, MGAL/B
3.0		ELECTRICITY, KWH/B
0.0183		FUEL, FOE

HYDROGEN TREAT HEAVY COKER GAS OIL FCCU FEED

FEED CONSTR.	PRODUCT STREAMS	
1.0		650-950 COKER GAS OIL
0.625		HYDROGEN, MSCF/B
	4.8	H2S AS SULFUR,LBS/B
	0.0332	FUEL GAS, FOE
	0.0302	N-BUTANE
	0.044	GASOLINE, 48 API
	0.266	DIESEL, 30 API
	0.700	FCCU FEED TREATED COKER G O
0.0734		VOLUME GAIN
1.0		CAPACITY CONSTRAINT FACTOR
0.017		STEAM MLBS/B
0.50		COOLING WATER, MGAL/B
3.0		ELECTRICITY, KWH/B
0.0183		FUEL, FOE

FLUID CATALYTIC CRACKING UNIT - PRETREATED FEED

CATALYTIC CRACKING OF PRE-TREATED 600-950 GAS OIL

FEED CONSTR.	PRODUCT STREAMS	
1.0		FCCU FEED, 600-950 TREATED
0.01887		FCCU REGEN COKE,MLBS/B
	0.146	H2S, LBS SULFUR/B
	0.0793	FUEL GAS FOE B/B
	0.0953	PROPYLENE
	0.0670	ISO-BUTANE
	0.0130	N-BUTANE
	0.0830	BUTYLENES
	0.3760	400 EP GASOLINE
	0.2380	LT CYCLE OIL-DESULFURIZED
	0.1220	HVY CYCLE OIL-DESULFURIZED
	0.0200	DECANT OIL - DESULFURIZED
0.0936		VOLUME GAIN
0.65703		CAPACITY CONSTRAINT FACTOR
	0.0566	STEAM PRODUCTION, MLBS/B
0.650		COOLING WATER, MGAL/B
0.195		ELECTRICITY, KWH/B
0.00193		FUEL FOE B/B

CATALYTIC CRACKING OF PRE-TREATED HEAVY COKER GAS OIL IN BLEND WITH 600-950 VIRGIN GAS OIL

FEED CONSTR.	PRODUCT STREAMS	
0.1272		FCCU FEED, TREATED COKER GO
0.8728		FCCU FEED, 600-950 TREATED
0.01983		FCCU REGEN COKE, MLBS/B
	0.446	H2S, LBS SULFUR/B
	0.0783	FUEL GAS FOE B/B
	0.0954	PROPYLENE
	0.064	ISO-BUTANE
	0.013	N-BUTANE
	0.082	BUTYLENES
	0.375	400 EP GASOLINE
	0.238	LT CYCLE OIL-DESULFURIZED
	0.122	HVY CYCLE OIL-DESULFURIZED
	0.02	DECANT OIL
0.0877		VOLUME GAIN
0.69046		CAPACITY CONSTRAINT FACTOR
	0.0616	STEAM PRODUCTION, MLBS/B
0.655		COOLING WATER , MGAL/B
0.207		ELECTRICITY, KWH/B
0.00197		FUEL FOE B/B

CATALYTIC CRACKING OF PRE-TREATED BLEND OF PDA GAS OIL AND 600-950 VIRGIN GAS OIL

FEED CONSTR.	PRODUCT STREAMS	
0.3531		FCCU FEED,TRTD PDA GAS OIL
0.6469		FCCU FEED,600-950 TREATED
0.02315		FCCU REGEN COKE, MLBS/B
	0.466	H2S, LBS SULFUR/B
	0.0842	FUEL GAS FOE B/B
	0.0924	PROPYLENE
	0.053	ISO-BUTANE
	0.012	N-BUTANE
	0.083	BUTYLENES
	0.372	400 EP GASOLINE
	0.261	LT CYCLE OIL-DESULFURIZED
	0.099	HVY CYCLE OIL-DESULFURIZED
	0.02	DECANT OIL - DESULFURIZED
0.0766		VOLUME GAIN
0.80606		CAPACITY CONSTRAINT FACTOR
	0.0701	STEAM PRODUCTION - MLBS/B
0.734		COOLING WATER, MGAL/B
0.277		ELECTRICITY, KWH/B
0.00256		FUEL FOE B/B

CATALYTIC CRACKING OF PRE-TREATED 700-950 VIRGIN GAS OIL

FEED CONSTR.	PRODUCT STREAMS	
1.0		FCCU FEED, 700-950 TREATED
0.02136		FCCU REGEN COKE, MLBS/B
	0.476	H2S LBS SULFUR/B
	0.0779	FUEL GAS FOE B/B
	0.0925	PROPYLENE
	0.0590	ISO-BUTANE
	0.0130	N-BUTANE
	0.0800	BUTYLENES
	0.3830	400 EP GASOLINE
	0.1930	LT CYCLE OIL-DESULFURIZED
	0.1670	HVY CYCLE OIL-DESULFURIZED
	0.0200	DECANT OIL - DESULFURIZED
0.0854		VOLUME GAIN
0.74373		CAPACITY CONSTRAINT FACTOR
	0.0426	STEAM PRODUCTION, MLBS/B
0.648		COOLING WATER, MGAL/B
0.225		ELECTRICITY, KWH/B
0.00219		FUEL FOE B/B

INCREASE FCCU CONVERSION

INCREASE FCCU CONVERSION FOR PRETREATED STOCKS VALUES ARE SCALED BY 100 FOR CONVENIENCE

FEED CONSTR.	PRODUCT STREAMS	
0.0510		COKE
	0.159	FUEL GAS FOE
	0.156	PROPYLENE
	0.25	BUTYLENES
	0.547	400 EP GASOLINE
	0.528	LCO FROM DESULFURIZED FEED
1.53		HCO FROM DESULFURIZED FEED
0.110		VOLUME GAIN
1.77577		CAPACITY CONSTRAINT FACTOR
	0.264	STEAM PRODUCTION, MLBS/B
0.53		COOLING WATER, MGALS/B
0.644		ELECTRICITY, KWH/B
0.0052		FUEL CONSUMED, FOE/B

INCREASE FCCU CONVERSION SCALED BY 100 FOR CONVENIENCE

FEED CONSTR.	PRODUCT STREAMS	
0.0510		COKE
	0.159	FUEL GAS,FOE
	0.156	PROPYLENE
	0.25	BUTYLENE
	0.547	400 EP GASOLINE
	0.528	LT CYCLE OIL
1.53		HEAVYCYCLE OIL
0.110		VOLUME GAIN
1.77577		CAPACITY CONSTRAINT FACTOR
	0.264	STEAM PRODUCTION
0.53		COOLING WATER
0.644		ELECTRICITY, KWH/B
0.0052		FUEL, FOE

BUTYLENE ALKYLATION

FEED CONSTR.	PRODUCT STREAMS	
0.585		FCCU BUTYLENES
0.690		ISO-BUTANE
	1.00	MOTOR ALKYLATE
	0.275	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
3.75		ELECTRICITY, KWH/B
0.17		PROCESS FUEL, FOE/B
0.011		STEAM, MSCF/B
3.68		COOLING WATER, MGAL/B

PROPYLENE ALKYLATION UNIT

FEED CONSTR.	PRODUCT STREAMS	
0.706		PROPYLENE
0.723		ISO-BUTANE
	1.0	MOTOR ALKYLATE
	0.087	PROPYLENE TO FUEL FOE B/B
	0.342	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
3.75		ELECTRICITY, KWH/B
0.17		PROCESS FUEL, FOE B/B
0.011		STEAM, MLBS/B
3.68		COOLING WATER, MGAL/B

CATALYTIC POLYMERIZATION

FEED CONSTR.	PRODUCT STREAMS	
1.2		PROPYLENE
	1.00	POLY GASOLINE
	0.20	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
1.7		ELECTRICITY, KWH/B
0.200		STEAM, MLBS/B
0.630		COOLING WATER, MGAL/B

HYDROGEN & SULFUR PLANT

HYDROGEN PLANT

FEED CONSTR.	PRODUCT STREAMS	
0.07870		FUEL GAS CONSUMED, FOE
	1.0	HYDROGEN, MSCF
	0.0787	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
	0.00432	STEAM, MLBS/MSCF
0.028		COOLING WATER,MGAL/MSCF
3.34		ELECTRICITY,KWH/ MSCF

AMINE PLANT
H2S REMOVAL BY MEA ABSORPTION

FEED CONSTR.	PRODUCT STREAMS	
1.0		H2S, SULFUR LBS
	1.0	REGENERATED H2S, LBS S/LB
0.005		STEAM MLBS/LB
0.015		ELECTRICITY KWH/LB
0.016		COOLING WATER MGAL/LB

SULFUR PLANT

FEED CONSTR.	PRODUCT STREAMS	
1.0		H2S,REGEN. AS SULFUR,LBS
	1.0	SULFUR, LBS
1.0		CAPACITY CONSTRAINT FACTOR
	0.00336	STEAM PRODUCED, MLBS/LB
0.050		ELECTRICITY KWH/B

UTILITIES

COOLING WATER UTILITY MGAL

FEED CONSTR.	PRODUCT STREAMS	
	1.0	COOLING WATER, MGALS
0.820		ELECTRICITY, KWH/MGAL

ELECTRIC POWER GENERATION

FEED CONSTR.	PRODUCT STREAMS	
	1.0	ELECTRICITY, KWH
0.00145		FUEL OIL, FOE

GENERATE STEAM FROM SALEABLE FUEL

FEED CONSTR.	PRODUCT STREAMS	
1.00		FUEL, FOE
	4.4	PROCESS STEAM, M LBS

MISCELLANEOUS REFINERY TRANSFERS

BURN RFFINERY FUEL GAS

FEED CONSTR.	PRODUCT STREAMS	
1.0		FUEL GAS
	1.0	PROCESS FUEL
	1.0	VOLUME LOSS

BURN PRODUCT FUEL IN REFINERY

FEED CONSTR.	PRODUCT STREAMS	
	1.0	HEAVY FUEL OIL PRODUCT
	1.0	PROCESS FUEL
	1.0	VOLUME LOSS

FLARE EXCESS REFINERY FUEL

FEED CONSTR.	PRODUCT STREAMS	
1.0		REFINERY FUEL FOE
	1.0	EXCESS REFINERY FUEL FLARED

BURN EXCESS HYDROGEN

FEED CONSTR.	PRODUCT STREAMS	
1.0		HYDROGEN
	0.053	FUEL GAS, FOE EQUIVALENT
0.053		VOLUME GAIN

BURN PROPYLENE

FEED CONSTR.	PRODUCT STREAMS	
1.0		PROPYLENE
	0.615	REFINERY FUEL
	1.0	VOLUME LOSS

BURN BUTANES

FEED CONSTR.	PRODUCT STREAMS	
1.0		NORMAL BUTANE
	0.702	REFINERY FUEL
	1.0	VOLUME LOSS

BURN ISO-BUTANE

FEED CONSTR.	PRODUCT STREAMS	
1.0		BURN ISO-BUTANE
	0.702	ISO-BUTANE FOE
	1.0	VOLUME LOSS

BUY ISO-BUTANE

FEED CONSTR.	PRODUCT STREAMS	
	1.0	BUY ISO-BUTANE
1.0		VOLUME GAIN

SUPPLEMENTARY DATA FOR KUWAIT OPERATIONS

HYDRODESULFURIZATION OF KUWAIT RESIDUUM – SINGLE PASS OPERATION

FEED CONSTR.	PRODUCT STREAMS	
1.0		KUWAIT VACUUM RESID
	0.0286	FUEL GAS$ FOE/B
	0.0050	ISOBUTANE
	0.0050	NORMAL BUTANE
	0.0500	REFORMER FEED (KUWAIT)
	0.3850	380-650 DES. KUW DISTILLATE
	0.3700	•650-1000 GAS OIL, LOW SULF.
	0.2710	DESULF. RESID.
0.1146		VOLUME GAIN
1.0		CAPACITY CONSTRAINT FACTOR
1.4540		HYDROGEN, SCF/B
0.0312		STEAM, MLBS/B
0.9650		COOLING WATER, MGAL/B
8.4000		ELECTRICITY, KWH/B
0.0218		FUEL OIL, FOE B/B

HYDRODESULFURIZATION OF KUWAIT RESIDUUM – RECYCLE TO EXTINCTION

FEED CONSTR.	PRODUCT STREAMS	
1.0		KUWAIT VACUUM RESID
	0.0390	FUEL GAS, FOE/B
	0.0070	ISOBUTANE
	0.0070	NORMAL BUTANE
	0.5270	380-650 DESULF. KUWAIT D.
	0.0690	REFORMER FEED (KUWAIT)
	0.5070	•650-1000 GAS OIL, LOW SULF.
0.1560		VOLUME GAIN
1.3702		CAPACITY CONSTRAINT FACTOR
1.9840		HYDROGEN, SCF/B
0.0428		STEAM,MLBS/B
1.3220		COOLING WATER, MGALS/B
11.5000		ELECTRICITY, KWH/B
0.0299		FUEL OIL, FOE/B

CAT CRACK KUWAIT GAS OIL

FEED CONSTR.	PRODUCT STREAMS	
1.0000		KUWAIT GAS OIL
	0.0580	FUEL GAS, FOE/B
	0.0500	PROPYLENE
	0.0530	ISOBUTANE
	0.0200	N-BUTANE
	0.0790	BUTYLENE
	0.5040	FCCU GASOLINE
	0.2540	LIGHT CYCLE OIL, KUWAIT
	0.0500	HEAVY CYCLE OIL, KUWAIT
	4.0000	SULFUR IN H2S, LBS/B
0.0260		COKE YIELD
0.0680		VOLUME GAIN
0.9150		CAPACITY CONSTRAINT FACTOR
	0.0980	STEAM PRODUCED, MLBS/B
0.8800		COOLING WATER, MGAL/B
0.2840		ELECTRICITY, KWH/B
0.0027		PROCESS FUEL, FOE B/B
0.5040		OCTANE ADJUSTMENT, KUWAIT

KUWAIT CRUDE UNIT OPERATION

FEED CONSTR.	PRODUCT STREAMS	
1.0000		KUWAIT CRUDE
	0.0048	GAS, FOE
	0.0040	ISOBUTANE
	0.0160	NORMAL BUTANE
	0.0370	KUWAIT C5 -IC6 LT GASOLINE
	0.2050	KUWAIT C6- 380 WIDE NAPHTHA
	0.1650	380-560 LT. DIST. KUWAIT
	0.0670	560-650 HVY DIST. KUWAIT
	0.3120	670-1100 KUWAIT FCCU FEED
	0.1860	VACUUM RESIDUE, KUWAIT
	0.0032	VOLUME LOSS
1.0000		CAPACITY CONSTRAINT FACTOR
0.6000		WATER, MGAL/B
0.0250		STEAM, MLBS/B
0.0130		FUEL, FOE
0.4000		ELECTRICTY, KWH/B
	0.0940	SULFUR ADJ. NAPHTHA PRTRT

SUPPLEMENTARY DATA FOR KUWAIT OPERATIONS

PRETREAT 400 EP NAPHTHA KUWAIT
H2S IN CRUDE + COKER MODELS

FEED CONSTR.	PRODUCT STREAMS	
1.0000		KUWAIT 275-380 EP NAPHTHA
0.0700		HYDROGEN
	0.9950	TREATED KUWAIT REF. FEED
	0.0050	VOLUME LOSS
1.0000		CAPACITY CONSTRAINT FACTOR
0.0130		REFINERY FUEL, FOE/B
0.7500		ELECTRICITY, KWH/B

REFORM KUWAIT NAPHTHA, 90 OCTANE

FEED CONSTR.	PRODUCT STREAMS	
1.0000		TREATED KUWAIT NAPHTHA
	0.1100	FUEL GAS, FOE/B
	0.0080	ISO-BUTANE
	0.0120	N-BUTANE
	0.8560	KUWAIT REFORMATE, 90 RON
	0.5180	HYDROGEN, SCF/B
	0.0140	VOLUME LOSS
1.0000		CAPACITY CONSTRAINT FACTOR
0.0647		REFINERY FUEL, FOE/B
0.0010		STEAM, MLBS/B
3.8000		ELECTRICITY, KWH/B
0.7450		WATER, MGAL/B

COKE KUWAIT RESID

FEED CONSTR.	PRODUCT STREAMS	
1.0000		KUWAIT VACUUM RESIDUUM
	0.0708	FUEL GAS
	0.0240	PROPYLENE
	0.0050	ISO-BUTANE
	0.0100	N-BUTANE
	0.0130	BUTYLENES
	0.0600	COKER LT GASOLINE
	0.1870	REFORMER FEED, KUWAIT COKER
	0.2560	LIGHT GAS OIL/FURNACE OIL
	0.2150	COKER GAS OIL TO FCCU
	4.6500	SULFUR IN H2S, LBS/B
	0.1156	BYPRODUCT COKE, MLBS/B
	0.1592	VOLUME LOSS
1.0		CAPACITY CONSTRAINT FACTOR
0.0380		STEAM CONSUMED, MLBS/B
1.4400		ELECTRICITY, KWH/B
0.0329		REFINERY FUEL, FOE/B
1.3000		COLLING WATER, MGAL/B
	0.4600	SULFUR ADJUSTMENT, REF FEED
0.0060		HYDROGEN ADJ. REF PRTRT.

INCREMENTAL FCCU YIELDS, COKER G.O.
FROM KUWAIT CRUDE

FEED CONSTR.	PRODUCT STREAMS	
1.0000		COKER HEAVY G.O. KUWAIT
0.0333		COKE ON CATALYST, MLBS/B
	0.1040	FUEL GAS, FOE/B
	0.1020	PROPYLENE
	0.0860	BUTYLENES
	0.3570	FUEL GASOLINE
	0.3450	KUWAIT CYCLE OIL
	6.0000	H2S AS SULFUR, LBS/B
	0.0640	ISO-BUTANE
	0.0130	N-BUTANE
0.0710		VOLUME GAIN
1.1600		CAPACITY CONSTRAINT FACTOR
	0.1720	
0.3460		
0.4340		
0.0034		

SUPPLEMENTARY DATA FOR KUWAIT OPERATIONS

CATALYTICALLY CRACK GAS OIL FROM RESID HYDRODESULFURIZATION OF KUWAIT VACUUM PITCH

FEED CONSTR.	PRODUCT STREAMS	
1.0		LOW SULFUR KUWAIT GAS OIL
0.02136		FCCU REGEN COKE, MLBS/B
	0.476	H2S LBS SULFUR/B
	0.0779	FUEL GAS FOE B/B
	0.0925	PROPYLENE
	0.0590	ISO-BUTANE
	0.0130	N-BUTANE
	0.0800	BUTYLENES
	0.3830	400 EP GASOLINE
	0.1930	LT CYCLE OIL-DESULFURIZED
	0.1670	HVY CYCLE OIL-DESULFURIZED
	0.0200	DECANT OIL - DESULFURIZED
0.0854		VOLUME GAIN
0.74373		CAPACITY CONSTRAINT FACTOR
	0.0426	STEAM PRODUCTION, MLBS/B
0.648		COOLING WATER, MGAL/B
0.225		ELECTRICITY, KWH/B
0.00219		FUEL FOE B/B

FCCU 800-950 GAS OIL

FEED CONSTR.	PRODUCT STREAMS	
1.0		FCCU 800-950 GAS OIL
0.03121		FCCU REGEN COKE MLBS/BBL
	3.47	H2S, LBS SULFUR/BBL
	0.0822	FUEL GAS,FOE
	0.0807	PROPYLENE
	0.0400	ISO-BUTANE
	0.0100	N -BUTANE
	0.0810	BUTYLENE
	0.3960	400 EP GASOLINE
	0.277	LCO
	0.083	HCO
	0.020	DECANT OIL
0.0699		VOLUME GAIN
1.13670		CAPACITY CONSTRAINT FACTOR
	0.09420	STEAM PRODUCTION, MLBS/B
0.96000		COOLING WATER, MGAL/B
0.340		ELECTRICITY, KWH/B
0.00318		FUEL FOE B/B

DESULF. OF IMPORTED RESIDUAL FUEL

FEED CONSTR.	PRODUCT STREAMS	
1.0		IMPORTED RESIDUAL FUEL OIL
	0.0236	FUEL GAS, FOE/B
	0.0034	I- BUTANE
	0.0047	N- BUTANE
	0.5748	380-650 DESULF KUWAIT DIST.
	0.0471	REFORMER FEED (KUWAIT)
	0.2493	•650-1000 GAS OIL, LOW FULF.
	0.1826	DESULF RESID
0.0855		VOLUME GAIN
1.1240		HYDROGEN, SCF/B
1.0		CAPACITY CONSTRAINT FACTOR
0.0266		STEAM, MLBS/BR
0.7579		COOLING WATER, MLBS/B
6.26		ELECTRICITY, KWH/B
0.0207		FUEL OIL, FOE B/B
	1.0	BUY FUEL GAS FOR H2 PLT

GLOSSARY

Alkylation Unit

A type of refining unit which produces very high quality gasoline from light olefins and isobutane. It is generally associated with FCCU's or delayed cokers which tend to produce excess light olefins.

Barrel: 42 U.S. gallons

BPSD: Barrels per stream day

Capital Charges: The portion of the operating cost which is directly proportional to the capital cost. Depreciation, insurance, total taxes, and cost of maintenance are typical examples.

Crude Unit: A refining unit comprising an atmospheric or low pressure first stage distillation unit, usually followed by a second stage distillation unit operating at vacuum. The unit comprises the distillation towers, heat exchangers and heaters.

Coker: A refining unit which is used to convert vacuum residuum to coke and distillates. A delayed coker is one type of coking unit.

Distillate: A refined or semi-refined material obtained by condensing the portion of a mixture which is vaporized when heated.

Distillate, Middle: The portion of petroleum boiling between 330°F and about 700°F. This material usually includes the stocks blended into No. 1 and No. 2 Fuel Oil.

FCCU: Fluid Catalytic Cracking Unit. This is one of the types of units used for converting high-boiling hydrocarbons into lower boiling hydrocarbons. These units use a finely divided catalyst which is conveyed from vessel to vessel in an aerated state. The aerated solid behaves like a fluid.

FCCU Feed Pretreater: A unit for pre-refining the feed stock charged to an FCCU. These units usually involve hydrogenation. After hydrogenation, sulfur and metallic compounds present in FCCU feed are practically eliminated and nitrogen is reduced. Such purification improves the operation of an FCCU.

Fuel, Distillate: No. 1 or No. 2 Fuel Oil

Fuel, Residual: Any liquid fuel containing the residuum from crude distillation or thermal cracking. In the context of this report, No. 6 Fuel Oil.

Gas Oil: While this term includes middle distillate, its more specific use is to designate the heaviest vaporizable portion of petroleum, ordinarily boiling from about 550 F to 1000 F or more. It is FCCU Feed.

Hydrocracking Unit: A recently developed process for cracking heavy hydrocarbons to light products in the presence of rather high partial pressures of hydrogen, and of a special catalyst. These processes can convert gas oil completely into gasoline and lighter fractions, or they can convert gas oils into high grade middle distillates. All hydrocracked products may be assumed to be essentially sulfur-free.

Hydrogen Plant: A unit for manufacturing hydrogen. In this study, the steam reformer type of hydrogen plant has been assumed throughout.

Hydrogen or H_2 Treater: A hydrogenating plant, often used to desulfurize or otherwise purify hydrocarbons.

Incremental Cost: The difference in cost of operation between some case, usually designated as a base case, and a related case involving one or more variants.

Isobutane: A specific hydrocarbon compound containing four carbon and ten hydrogen atoms. Isobutane reacts with light olefins in the presence of sulfuric or hydrofluoric acids to make high octane alkylate gasoline.

Iteration: A calculation which is repeatedly made in the course of solving the matrix representing an LP problem.

LP: Linear Programming. A mathematical operation, used for arriving at an optimal solution in a system where a number of operable solutions are possible. As used in the context of this study, the technique is used to seek out the most profitable method of either building a new refinery or operating or modernizing an existing refinery.

Matrix: An array of numbers, usually the coefficients of the variables used in the mathematical model of the refinery. The LP computer code recognizes the significance of these numbers, and manipulates them to arrive at the optimal solution for a given problem.

Octane: An abbreviation for octane rating. The octane rating is a means of ranking gasolines according to their resistance to detonating explosions when used as fuels in internal combustion engines. The scale is based on 0 for normal heptane and 100 for 2-2-3 tri-methyl pentane, often called iso-octane. Gasolines can also be rated above 100 octane by means of an extrapolation formula.

Olefin: A hydrocarbon which contains less atoms of hydrogen than its full complement; olefins are particularly characterized by one double bond between two carbon atoms. The double bond imparts characteristic chemical behavior.

Propane De-Asphalter: A process which extracts FCCU feed from residuum by its solubility in liquid propane.

Realization: A gross profit calculated by subtracting variable charges that are affected by processing rate from value of products produced. It is a convenient measure of profitability.

Reformer: A catalytic refining unit, usually employing a platinum-containing catalyst. The reformer upgrades low octane gasolines into high octane gasolines by rearrangement of the molecular structure. Reformers generally yield by-product hydrogen.

Reformer Pretreater: A hydrogenation unit used for purifying the feed to reformers. The pretreater in effect protects the investment in platinum catalyst by insuring removal of catalyst poisons.

Residuum: The most non-volatile portion of petroleum, residuums are sometimes called long-or short-residuums. A long residuum is the non-volatile residue from an atmospheric pressure distillation; whereas a short residuum is obtained from vacuum distillation.

Resid HDS Unit: A residuum hydro desulfurizer. In actual fact, these units commonly also hydrocrack residuum. In any case, they operate in the presence of a special catalyst and a fairly high hydrogen partial pressure. Hydrogen is consumed in the process and sulfur is released as H_2S.

Shadow Price: This is the mathematical quantity used by the LP as a measure of profitability. It may be described as the cost of making the last barrel of a particular product. If this cost is below the value for that product, the profitability could be increased by making more.

Stream Day: A day on which a unit is operated at full capacity for 24 hours. Depending on the amount of maintenance work required, a unit may accumulate 300 to 330 or even more stream days in a year.

Sulfur Plant: As used in this report, a plant for converting H_2S to sulfur, using the Claus process.

BIBLIOGRAPHY

Much of the information used in this study was obtained from private unpublished sources. Included in this category are the assay data from which crude unit yields were adopted as well as other yield and cost data used in the formulation of the LP matrix. Even where published information was used, it was adapted to fit the requirements of this study. For these reasons, complete documentation of all sources of data is not practical.

The following is a list of selected references which may be of interest to the reader. Those references specifically referred to in the text of this report are noted under the section heading indicated.

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Delayed Coking Process

Mekler, V. and M. E. Brooks, "New Developments in Techniques in Delayed Coking," Proceedings, American Petroleum Institute, 1959, Section III, Refining, American Petroleum Institute, New York, New York, pp. 229 - 245.

Comparison of Fuel Oil Reduction Processes

The following paper by Slyngstad and Feigelman was used as a basis for initiating this study. The paper was published in abridged form by the Oil and Gas Journal.

Slyngstad, C. E., & Feigelman, S., "Fuel Oil Reduction," Paper presented at 38th Annual Meeting, California Natural Gasoline Association, Anaheim, California, October 10 - 11, 1963.

Oil & Gas Journal, "Four Roads Offered to Less Fuel Oil," Jan. 6, 1964, pp. 65-68.

General Refinery Processing Methods

Refining Process Handbook Issue 1962, Hydrocarbon Processing and Petroleum Refinery, Volume 41, No. 9, (September 1962).

Nelson, W. L., Petroleum Refinery Engineering, 4th ed. McGraw Hill Book Company, Inc., New York, New York (1958).

CRUDE SELECTION AND PRODUCT SLATE

Manne's Jet Fuel Study

Manne's LP study for jet fuel illustrates how a large scale LP simulation is set up. This reference also includes other economic simulation studies and provides useful background material.

Manne, Alan S., and Markowitz, Harry M., "Studies in Process Analysis - Economy Wide Production Capabilities," Cowles Foundation Monograph 18, John Wiley and Sons, New York, New York (1963). See Chapter 4 particularly.

California Crude & Products Data

Slyngstad, C. E., and Feigelman, S. (Reference listed above).

Brown, C. T., "California Crudes Show Great Variation in Properties," The Petroleum Engineering, Vol. 28, n. 1, pp. C 9-C 14, Jan. 1956.

FORMULATING THE MATHEMATICAL MODEL

There appears to be no completely satisfactory reference dealing with the technique of LP model formulation as related to the petroleum industry. The book by Manne and Markowitz referred to above is helpful as well as the following article.

Cheveny, John E., three part article appearing in Oil and Gas Journal:
Part 1: "What is Linear Programming?" Vol. 58, No. 10, pp. 113-116, 119-120 (March 7, 1960).
Part 2: "Putting Linear Programming to Work," Vol. 58, No. 12, pp. 108-110 (March 21, 1960).
Part 3: "Putting Linear Programming to Work," Vol. 58, No. 14, pp. 114-115, 117-118 (April 4, 1960).

Formulation of the matrix is to some extent related to the particular algorithm and computer program used. This study employed a proprietary program available from:

Bonner and Moore Associates, Inc.
500 Jefferson
Houston 3, Texas

Low cost natural gas fuel was considered in one LP case. Information on the long term availability of gas and predicted use patterns for it and other fuels is covered in two Federal Power Commission reports:

Federal Power Commission, National Power Survey, Advistory Com-Committee Report No. 21 on Fuels for Electric Generation (December 1963).

----, Advisory Committee Report No. 18 on Fuels for Electric Generation in Western United States. (July 1963)

HYDROGEN BALANCE

Utilization of byproduct hydrogen in a refinery is described in detail by Gwin.

Gwin, G. T., "Optimal Utilization of Byproduct Hydrogen in an integrated Oil Refinery," Proc. A.P.I., 1959, Section III, Refining, pp. 193-201.